

TROJAN POWDER COMPANY

ENGINEERING

DEPT. FILES

AT 100

EVAPORATION

BY

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The MODERN LIBRARY of CHEMICAL ENGINEERING

So rapid has been the development of the technology of chemical processes during the past decade it is only natural that literature on the subject has failed to keep pace.

Those in a position to write authoritative books have been too busy solving practical problems to chronicle their experiences for others, and the feeling that chemical engineering data were still coming in at a rapid rate caused many to hesitate to publish data which might soon be rendered out of date by further advances in practice.

During the past two years, however, it has become generally recognized that the best interests of the industry and the profession demand a literature dealing with the fundamental processes of chemical engineering, not only as a basis for future progress but as an economical factor in current operation, as well as for the instruction of the younger men entering the profession.

Such a literature Germany has long had, and any American who has worked or studied in Germany knows what a great factor it has been in Germany's chemical development. It is not believed, however, that any good purpose could be served by a general translation of the German books, as much of the equipment discussed therein never has been and never will be used in this country. As a matter of fact, much of this German literature is now out of date, not only in this country but abroad, and in many instances our technical knowledge of chemical processes outstrips the world.

In view of these conditions, The Chemical Catalog Company, Inc., has undertaken the work of interesting experts of indisputable standing in their several lines, in writing a series of technological works dealing with such fundamental processes in chemical engineering as Evaporation, Filtration, Distillation, Heat Transfer, Drying, Transportation of Liquids, Compression of Gases, Mechanical Handling of Materials, Grinding, Pulverizing, etc.

The scope of this Series naturally cannot be fixed at this time, but books in this group, known as the "Modern Library of Chemical En-

gineering" will develop as chemical engineering knowledge develops. An effort will be made to drag forth facts and little tricks of the trade from their hiding places so that the current status of all the fundamental subdivisions of chemical engineering can be presented without reserve. Where this is not possible through the agency of a single author or group of authors, the coöperation of all the owners of facts will be sought to that end, and an editor or board of editors will be appointed to handle existing data with fairness to all.

The Modern Library of Chemical Engineering series is uniform in format and binding. Some of the subjects, with their authors, in addition to this volume on "Evaporation" by Alfred L. Webre and Clark S. Robinson, are as follows:—

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By Arthur Wright

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Published January, 1926. Price \$5.00.

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In Preparation.

Announcements regarding the progress of the Modern Library of Chemical Engineering will be made from time to time, and of course more detailed information will be furnished at any time upon request. The publishers will welcome suggestions that will help improve the service these books on the technology of chemical processes are expected to render to chemical engineers, plant operatives, research men and students.

The CHEMICAL CATALOG COMPANY, Inc.

THIS BOOK IS RESPECTFULLY DEDICATED

TO MY GOOD FRIEND

MR. JOHN H. MURPHY

FROM WHOM I LEARNED THAT

NOTHING WORTHWHILE CAN BE ACCOMPLISHED

WITHOUT HARD WORK AND COMMON SENSE

ALFRED L. WEBRE.

Preface.

This book was written at the request of the Chemical Catalog Company, Inc., and contains theoretical and practical data accumulated during some twenty years of actual experience. Its object is to present more recent conceptions of the design and operation of evaporators.

An attempt has been made to arrange the subject matter in such a way as to appeal to the average industrial engineer, and the discussions have been couched in language that should be easily understandable.

Free use has been made of illustrations, as it is felt that these contribute materially to the clarity of the text.

It is therefore hoped that the volume will constitute a contribution of value to industry.

The author wishes to thank various manufacturers for the generous assistance he has received in the laborious preparation of this work, notably in the fourth section of the book in which specific acknowledgments are made.

Many thanks are also due to the following firms and individuals:

Mr. John H. Murphy, of New Orleans, for the valuable practical information secured from him during many years of friendly co-operation;

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ALFRED L. WEBRE.

Contents.

	PAGE
INTRODUCTION—What the Owner of Every Evaporator Should Know . . .	9

SECTION I

Theoretical Considerations.

CHAPTER		
1.	EVAPORATION	12
2.	VAPORIZATION AND VAPOR PRESSURE RELATIONS	18
3.	TRANSMISSION OF HEAT	24
4.	CONDUCTIVITY OF SURFACE FILMS	32
5.	FACTORS AFFECTING THE COEFFICIENT OF HEAT TRANSMISSION	38
6.	BASIC PRINCIPLES	51
7.	THE HEAT BALANCE	56
8.	PRACTICAL USE OF THE HEAT BALANCE	69
9.	VAPOR HEATING	80
10.	VAPOR COMPRESSION	93
11.	VAPOR PIPES	96
12.	ENTRAINMENT AND ITS PREVENTION	101
13.	NATURAL CIRCULATION	109
14.	MECHANICAL CIRCULATION	119
15.	SELECTION OF TUBE PROPORTIONS	126
16.	TYPES OF STEAM BELTS	130
17.	SINGLE EFFECT FINISHING PANS	141
18.	EVAPORATOR CONDENSERS	144
19.	SPRAY PONDS AND COOLING TOWERS	155
20.	MULTICURRENT SOLUTION HEATERS	160

SECTION II

Information on the Operation of Evaporators

1.	INFORMATION ON THE OPERATION OF EVAPORATORS	171
2.	CHARTS FOR EVAPORATORS	172
3.	STARTING UP	178
4.	SHUTTING DOWN AND CLEANING	186
5.	SCALING AND FOULING	189
6.	CONDENSATE REMOVAL	193
7.	NON-CONDENSABLE GAS REMOVAL	200
8.	FEEDING SYSTEMS	203
9.	PRACTICAL OPERATION OF JET CONDENSERS	217

CHAPTER	PAGE
10. VACUUM PUMPS AND THEIR CHARACTERISTICS	223
11. EVAPORATOR ACCESSORIES	231
12. INSULATION	235
13. INVESTIGATING TROUBLES	237

SECTION III

Applications to Various Industries.

1. GENERAL APPLICATIONS TO SPECIFIC INDUSTRIES	243
2. THE HEAT BALANCE IN THE CANE SUGAR FACTORY	244
3. THE HEAT BALANCE IN THE BEET SUGAR FACTORY	293
4. EVAPORATION OF CRYSTALLIZING SOLUTIONS	345
5. EXTRACTS AND DYES	354
6. ORGANIC AND FOAMY MATERIALS	356
7. PAPER MILL WASTE LIQUORS	359
8. DELICATE MATERIALS	390
9. THE HEAVY GROUP	392
10. THE EVAPORATION OF GLYCERINE LIQUORS, AND THE WURSTER & SANGER EVAPORATOR	394

SECTION IV

Types of Evaporators.

1. INTRODUCTORY	407
2. THE STANDARD EFFECT	408
3. THE BAFFLED EVAPORATOR	412
4. THE HORIZONTAL TUBE EVAPORATOR	417
5. THE LILLIE EVAPORATOR	424
6. "BUFLOVAK" INCLINED RAPID CIRCULATION EVAPORATOR	427
7. THE KESTNER EVAPORATOR	434
8. THE BADGER HIGH SPEED EVAPORATOR	440
9. THE BASKET TYPE EVAPORATOR	443
10. THE GARRIGUE EVAPORATOR	447
11. THE MANISTEE EVAPORATOR	452
12. DISTILLED WATER EVAPORATORS	456
13. THE ZAREMBA PATENT EVAPORATOR	461
14. THE HIGH CONCENTRATOR	470

Introduction.

What the Owner of Every Evaporator Should Know.

An evaporator is a highly developed apparatus, based on a mass of theoretical and practical knowledge which cannot be ignored if efficient results are desired. It is not simply a boiling pot, and he who proceeds on any such assumption will have to pay the penalty of his folly.

The outstanding difficulty lies in the great number of variables to be taken into consideration, some independent, some interdependent. It has been estimated that there are at least fifty such variables, all affecting the final design. It is therefore manifest that this work should be entrusted to men of knowledge and experience whose training enables them properly to grasp, interpret, and solve with satisfaction, efficiency, and economy the problems encountered.

It has been said that anyone can accomplish any given task, if he has at his disposal unlimited time and financial resources, two factors always found in limited quantities. It takes knowledge and experience, however, to do the same task quickly and effectively at reasonable cost.

A well-designed apparatus will do its work without effort. It will cost considerably less than one improperly planned and in whose design established principles have been overlooked. It will also perform with greater economy of steam, which is always an important consideration.

There is no science in which there is more misinformation, nor where the rule of thumb has played a greater part, and there are many evaporators in use to-day whose performance could be materially improved by a few simple and inexpensive changes.

About twenty years ago, the author, then a draftsman with a sugar engineering firm, had the task of putting into execution plans for a multiple effect evaporating apparatus, and on one occasion, was told to make the vapor pipe between two adjacent bodies 23 inches. The engineer in charge would entertain no discussion on the subject, and the dimension was retained. It was beyond comprehension that the proportions of evaporators were so rigid and delicately balanced that a 24-inch standard sized pipe could not be used, and that such unusual sizes should prevail. On another occasion, this same engineer maintained that it was not practical to use multiple effects with more than three bodies. In view of present day practice, the lack of vision on the part of some of those in charge of this work was really startling.

Even to this day, there are on all sides examples of misunderstanding or lack of knowledge in some of the largest plants. For instance, there is no good reason why sugar factories in Cuba should burn any coal or oil whatever, for the fiber in the cane is ample, except under abnormal conditions, to supply all the steam requirements, if the heating and evaporating station is properly designed. As a matter of fact, they nearly all

burn large quantities of such additional fuel, some centrals burning as much as three to four gallons of oil per bag of sugar. In such a plant, producing 300,000 bags per annum, even with oil at four cents per gallon, this would amount to from \$36,000 to \$48,000 per annum, an amount probably sufficient to pay the cost of the apparatus required to effect the saving.

The value of steam savings in proportion to cost of operation becomes greater as the selling price of the ultimate product is low, and the cost of fuel high. For instance, steam economy is of secondary importance in the manufacture of photographic gelatine, the price per pound of which is high, but assumes extreme importance in the manufacture of tanning extract, the price per pound of which is relatively low, and for the production of which relatively large quantities of water have to be evaporated. Similarly, with a salt plant obtaining its fuel from saw mill refuse one is not interested in steam economies if there is an excess of wood which must be burnt to get rid of it, but becomes very much interested if coal must be bought at high prices instead. It is therefore necessary to consider the economics of the situation in appraising the value of heat savings.

In the study of evaporator problems, it is always advisable to consider at the same time the heating operations necessary in the plant, for, by combining the two, it is very often possible to secure additional steam economies which are well worth while, and which, in reality, do not involve unreasonable complications. Particularly in the beet sugar industry, these combinations have been elaborated with remarkable success, and to-day, no beet sugar man would put up an evaporator that is not designed to furnish juice vapors from the various bodies for heating operations. In many industries, similar applications are easily possible, but never used, mainly due to lack of information on the part of the owners and managers of the plant.

As to size of apparatus, it has been well established that ample evaporator capacity is always a good investment, and shortage in this respect is invariably accompanied with retrenchments in some other department of the plant, thus bringing about poor yields, and high fuel consumptions.

It is perhaps well to rectify a common error on the part of laymen as regards the performance of multiple effects. It is only a very crude and misleading approximation to say that a single effect evaporates one pound of water per pound of steam used, a double effect, two pounds of water per pound of steam, a triple, three pounds, and so on. This fact has been well demonstrated in the discussion of the heat balance, Chapter 7, to which the reader is referred.

It has been the purpose of the authors of this book to set down clearly, as far as they are able, their version of the information on the subject of evaporation as known to-day, in order that the reader in turn may secure a comprehensive grasp of this interesting branch of engineering.

There are four main divisions of the subject matter. The first comprises the theoretical considerations; the second, general practical information; the third, broad applications to several important industries; and the fourth, a discussion of the types of evaporators usually encountered.

SECTION I

THEORETICAL CONSIDERATIONS

In this, the first section, is given a general discussion of the theory of Heat Transmission, as well as the elaboration of the principles of Single and Multiple Effect Evaporation, and its corollaries.

Naturally, a considerable part of this data has been secured from outside sources, the idea being to give a comprehensive survey of the subject.

Much of this information, however, represents the author's own conceptions and methods of analysis which have been used for many years, and it is hoped that they may prove to be similarly useful to others.

Many of our existing theories are so incomplete that it is very difficult to make practical use of them. They have been presented to the best of our knowledge with the hope that they will illuminate some of the puzzling problems encountered.

In practice, it is necessary to have a working basis, and it has been attempted to give such a basis.

Some of the discussions will be found to depart considerably from the well-beaten paths in an attempt to look at the subject matter from different angles, in order to provide at least a logical explanation of observed facts.

Chapter I.

Evaporation.

Definition. Evaporation will be understood here as the physical process of concentrating solutions by removing part of the water from them by vaporization.

Physical Process. In the first place, it is a physical process, and there will be no chemical change except under very special conditions. The well-known laws of physics are therefore the ones to apply.

Removal of Water. Secondly, it is a process of concentrating solutions by the partial removal of water. A solution, in the sense used in this text, consists of water with some material dissolved in it. Unless otherwise specified, it may be assumed that this dissolved material is a solid, and not a liquid. The partial or total separation of mixed liquids is usually accomplished by an entirely different process called distillation, with which this text is not concerned.

The most convenient way to express the strength of solutions quantitatively is as "per cent. by weight of solids in solution."

Density Scales. There are many arbitrary scales to measure the concentration of solutions which have been in use in the various industries, and in the appendix will be found a list giving the scales, and their corresponding specific gravities. The readings are usually made by means of a hollow glass spindle called a hydrometer, weighted at the bottom, having an enlarged section above which is a long sealed tubular glass of uniform diameter whose length is calibrated to read the densities. The spindle floats in a vertical position, and the depth of the immersion on the tubular part is inversely proportional to the change of density.

Among the various scales in use may be mentioned the following:

1. The Baumé scale originally intended to represent the percentage of solids in a pure sodium chloride solution.
2. The Brix scale which is supposed to represent the percentage of solids in solution for pure sucrose.
3. The Twaddell scale, used in many industries such as the manufacture of tanning extract, is based upon the specific gravity of the solution, consisting of the first two digits on the right of the decimal point of the specific gravity expression multiplied by two, thus a specific gravity of 1.21 corresponds to a Twaddell reading of 42.
4. The Barkometer scale, used almost exclusively in the tanning extract industry, consisting of the first three digits on the right of the decimal point in the specific gravity expression, so that a specific gravity of 1.014 would correspond to a Barkometer reading of 14.

5. The Gravity scale, which gives a direct reading of the specific gravity of the solution.

The reading of each of these density scales is correct only at a certain fixed temperature, and if the solution is spindled, this temperature should be reached before the reading is made, otherwise it will be necessary to make a temperature correction to be added if the solution is above the standard temperature, and deducted if below. The established standard is usually 17.5°C. , or 63.5°F. , although this is not absolutely fixed, the temperature at which the spindle is calibrated usually being stamped thereon.

Percentage of Evaporation. In calculating the percentage of evaporation by weight required to concentrate from one density to another, it is best to calculate this by converting the densities into percentage of solids in solution by weight. Complete tables for such conversions can be found in chemical handbooks. The computation is quite simple, consisting merely of determining the ratio of the difference of the percentages to the final percentage. Thus evaporating from 15 per cent. solids to 60 per cent. solids is $\frac{60 - 15 = 45}{60} = .75$, or 75 per cent. evaporation by weight.

Saturated Solutions. The above is true only as long as the solution is below saturation, or the point at which it will begin to crystallize. When this point is passed, if there is only one salt in solution, the density will remain the same, and is therefore no longer a measure of the amount of evaporation. All additional evaporation will be accompanied by the precipitation of salt. A good example of this condition is in the manufacture of salt by the vacuum process from an initially saturated solution of pure sodium chloride.

The solubility always varies with the temperature of the solution, and although it usually increases with increasing temperatures, it varies greatly with different materials. Even in the same solution, it will not increase uniformly, but changes with the range. For certain materials it may increase up to a certain point, and decrease beyond it. A good example is sodium sulphate.

In solutions of mixed salts, with two, three, four, and even five different materials in solution, the problem becomes very complicated, as the salts affect each other's solubilities, and change their characteristics in this direction. This is really a study in itself, and is beyond the field of this book, belonging to that of physical chemistry.

Vaporization. At the present time, the only practical method of securing concentration of solutions has been found to be the actual conversion of the water into vapor. In the future, science may develop another more economical means of doing it, but so far, it is the cheapest available method and practically the only one.

Freezing. It has been found possible under certain very limited conditions to accomplish concentration by freezing out part of the water, but this process is at present so expensive that it is usually out of consideration. An illustration is the manufacture of raw water ice. If raw

water, containing salts in solution is frozen in a can, the pure water crystallizes at the periphery of the container, leaving the other material in the unfrozen core at the center. The freezing process is stopped at a certain point, the core, containing the solid material in a concentrated solution, poured out, and the empty space filled with distilled water. The freezing process is now resumed, and the resulting block of ice will be perfectly transparent, and nearly free from foreign matter.

Experiments have also shown that it is possible to concentrate solutions of weak acetic acid by freezing.

These examples indicate that evaporation is not the sole possible method for concentration, although it may now be the only practical one, and it is not beyond the bounds of possibility that the future may disclose some new and as yet unthought of process for accomplishing it.

Application of Heat. In order to vaporize, there must be an application of heat somewhere in the process. This may come either from natural sources, or from artificial sources by the combustion of fuel.

Solar Ponds. A good example of the use of natural heat for vaporization is the operation of solar ponds, which have been used for centuries throughout the world, usually for the concentration of brines in the production of common salt. Some of these ponds have very large capacities. There is one in the western part of the United States that evaporates 5000 gal. of water per minute. The brine does not boil in this case, but merely evaporates from the surface, the moisture being picked up by the air, much as water evaporates from the clothes hanging out in the family back yard. Atmospheric conditions determine the feasibility of these processes, and seasonal variations will therefore have a great bearing on the operation, varying between wide limits from summer to winter. High rates of evaporation will be obtained in dry, sunny, warm, windy days, and low rates in cloudy damp weather. The relative humidity of the air also makes a great difference. When the saturation point of the brine is reached, the evaporation slows down to only a fraction of what it was before, due to the fact that the surface of the ponds becomes partially covered with salt crystals floating on the brine, thus partially blanking off the evaporation area. In one case in India, this problem has been overcome by mechanically agitating the surface, thus causing the floating salt crystals to sink, eliminating the insulating effect.

No rule or formula can be given for this work, the rate of which must be determined experimentally in each case with great care. Usually, there are a number of ponds in series, the last, or crystallizing ponds being in parallel. The process is allowed to go on until a fair harvest of salt has accumulated, when the particular pond is drained, allowed to dry, and the salt is raked into piles for shipment.

The bittern, or final mother liquor, containing residual salts, has to be eventually discarded, as it builds up in impurities, which would otherwise contaminate the salt. This bittern is sometimes further evaporated for the recovery of valuable materials, such as bromides, which it may contain, and this supplemental process is usually carried out in a steam evaporating plant, where accurate temperature control can be obtained.

The operation of solar ponds requires a very large acreage of ground, as the evaporation per unit of area is relatively small, and this is doubly true if climatic conditions are not favorable. They are generally operated in cycles to correspond to the seasons of the year, one period being devoted to evaporation and crystallization, and the other to harvesting, storage and marketing.

The land used for the construction of solar ponds should have a clay bottom, otherwise there will be large leakage losses. Usually, low lands near the sea are used where the tides can be made to feed the brine into the system. In some cases, according to the conditions, it may be necessary to pump the brine, but usually the ponds are arranged to have gravity flow from one to the other, and the only labor required is that of a man to set the gates in the connecting conduits according to the variable rate of evaporation.

When conditions justify, sometimes sprays are used, thus exposing very large evaporating surfaces, and speeding up the work, thereby necessitating much smaller areas for a given amount of work. When saturation is reached, however, it becomes very difficult to keep the nozzles from choking with precipitated salt. The power expense is naturally quite an item of cost, and this method can only be used where there is cheap power. Furthermore, there is a well-defined limit to the amount that can be evaporated in this way in a restricted area, for the brine being sprayed very quickly comes to the wet bulb temperature of the air beyond which it is impossible to force the work.

Many projects have failed due to lack of consideration of this factor, the plant having been erected on the assumption of a direct relation to small scale experiments in which information cannot be expected to be correct on account of reflected heat from surrounding objects. It is therefore advisable to proceed with care and judgment, and not to jump at conclusions without thorough investigation.

Other Sources of Natural Heat. Another source of natural heat has been found in the hot geysers and steam wells. There are several cases where this source of heat is utilized, and one in Italy, in which steam from the ground at 45 lbs. gauge is actually used for power purposes on a turbine. These cases are so rare, however, that they need only a passing mention.

Heat from Burning Fuel. In the great majority of cases, no natural heat is available, and one must resort to the burning of fuel. Here, there are two possibilities, first in the accomplishment of the vaporization by bringing the products of combustion into direct contact with the solution being concentrated. This can be done in only a few limited cases, the most important one being in the concentration of sulphuric acid through an acid tower or in cascade type concentrators. The essential considerations are first, whether or not the solution in question will be affected adversely by either the composition or the temperature of the products of combustion of the fuel used, which contain the heat liberated. Very good results are obtainable by this simple means, when there are no objections. The other consideration is the economy from the standpoint of

the cost of the fuel, as this is "single effect" operation, where the heat is used for evaporation only once.

In such arrangements, the solution must be sprayed, or percolated through some form of tower so as to expose a large amount of surface, for the heat conductivity of the gases resulting from combustion is not great.

Heat Transmission through Surfaces. Generally speaking, however, the heat generated by the combustion of fuel is transmitted through surfaces, the simplest form being the direct fired kettle.

Direct Fired Evaporators. Here, the products of combustion are brought into contact with a tank, container or kettle containing the liquid to be concentrated, and the heat is transmitted through the walls of this container into its contents which evaporate by boiling. This is perhaps the oldest form of evaporation. To-day, it finds only very limited application on account of its inherent inefficiency. There might be mentioned among these, the final concentration of caustic soda and calcium chloride, whose viscosities, boiling points, and corrosive properties make them practically impossible of concentration in steam evaporators at such densities.

The efficiency of the open kettle is very low on account of its small heating surface, its low heat transmission, and its range of boiling temperature necessarily fixed by atmospheric pressure and for those reasons, it is not in wide use to-day. In the old days, in the sugar industry, the so-called open kettle process of concentration was the only one known, and here it was customary to use a battery of cast iron pots in a line of graduated sizes, the smallest pot being directly under the fire, and the largest the farthest from it. The juice was run counter current to the flue gases, and entered uncooked into the big pots farthest away from the fire, where it was limed, heated, skimmed, being afterwards transferred to the next one preceding by the use of large buckets mounted on wooden poles resting in oar locks between each pot. When the small pot directly under the fire was "cooked," its contents were quickly removed by an operator with a pole bucket and dumped into a little wooden tank car on wheels, which in turn emptied its contents into the wooden crystallizing vats in the "purgery." When the small pot was emptied, the next in line was transferred, and then the preceding and so on, thus starting a new batch.

Such operations were necessarily limited to small scale, and represent the sugar industry as it existed about fifty years ago. Even then, many such plants had been dismantled for more modern equipment, and no one would think to-day of using this or similar processes except in rural districts where very small operations are carried on.

Sometimes, there is a special reason for using direct fired evaporation. For instance, the Georgia cane syrup is all made in open pans by direct fire, the contention being that due to caramelization, or incipient burning of the sugar, an especially desirable flavor is imparted making the product more palatable, and therefore giving it a better market. The same may be said of the maple sugar industry.

Steam Evaporation. In ordinary cases, probably 95 per cent. of the evaporation is done by generating steam in a boiler, and afterwards accomplishing the evaporation by using this steam in an evaporating apparatus having coils, tubes, or some such form of heating surface.

In this way, it is possible to use the live steam in prime movers for the generation of power, after which at reduced pressure in the form of exhaust steam it will serve very well for the work in hand. Thus, the cost of power so generated is only the small mechanical equivalent of heat plus minor radiation losses. This gives almost free power, or power at such a small cost that it can almost be neglected.

According to the conditions imposed, within certain limits, it is possible to control the temperature at which evaporation takes place by carrying on the reaction in a closed body, and disposing of the vapors by condensing in a jet or surface condenser. Were this not possible, there would be to-day no white refined sugar which is actually crystallized from a solution while in ebullition under vacuum.

These limits of temperature are determined by the commercially obtainable vacuum on one side, and by the pressure and temperature of steam on the other. By the use of live steam, it is possible to evaporate at a higher temperature by putting a back pressure on the evaporating body. If the exhaust pressure is sufficient, in this case also, it is possible to evaporate under pressure. The pressure on the liquor side of the evaporator must be sensibly lower than that on the steam side, otherwise there would be no difference of temperature between the two, and hence no evaporation, since this difference is what causes the flow of heat and is a measure of the amount of heat transferred.

The use of steam for evaporation has had as a direct and logical consequence the development of multiple effect evaporation, by means of which many large industries of to-day have been able to grow.

These conditions and their applications in practice will be fully discussed in the pages that follow, as they are the theme of this book.

Chapter 2.

Vaporization and Vapor Pressure Relations.

Evaporation. Evaporation, as considered in this book, involves the vaporization of a volatile solvent, usually water, and its removal from a solution which is concentrated thereby.

Mechanism of Evaporation. According to modern theory, a volatile liquid consists of a great number of exceedingly small portions of matter known as molecules which are in a very rapid state of irregular vibratory motion. This motion is not definitely known, but it is supposed to be similar to that of certain other collections of matter which are called colloidal particles, and whose movements may be studied by means of the ultramicroscope. Such particles collide frequently with each other and rebound or are deflected in new directions with undiminished speed. Molecules on the surface of a liquid are therefore in rapid motion in all directions and when one of them chances to be propelled beyond the surface layer of the liquid at a speed sufficient to overcome the surface tension of the liquid which holds it back, it will continue its flight out into the vapor space above. After a number of these molecules have so escaped, collisions will take place in the vapor space, and molecules will begin to shoot back into the liquid.

Definition of a Saturated Vapor. If this interchange between the liquid and the vapor continues long enough, a time will come when the same number of molecules will pass back into the liquid as go out into the vapor space in a given time. This is a state of equilibrium and the vapor is then said to be saturated with respect to the liquid under the particular conditions.

Mechanism of Vapor Pressure. The motion of the molecules in the vapor state is such that if the vapor be unconfined, the molecules tend to gradually spread apart from each other in all directions until the vapor is wholly dissipated. If, however, they are confined in an enclosed space, they will beat against the enclosing walls in an attempt to get out, thus developing a pressure. This pressure is proportional to the number of molecules present in the vapor and is a function of the speed at which they move. It has been found that for any given volatile liquid or solution, at any given temperature, there will be a definite pressure developed when the liquid remains in contact with vapor. This pressure is known as the vapor pressure of the liquid at that temperature.

Rise of Pressure with Temperature. The speed of the molecules increases with rising temperature. This causes greater numbers to escape from the surface of the liquid and when equilibrium is established, there

will be more present in the vapor space. This increase in vapor concentration and the incidental increase in the speed of molecular vibration causes an increase in the vapor pressure. In other words, the vapor pressure increases with the temperature.

Relation between Pressure and Temperature. There is a thermodynamic relationship which may be developed between the vapor pressure and the temperature for any liquid but it is extremely complicated even in the simplest cases, and is largely of theoretical interest.* Certain approximations may be made, however, which give the following relation:

$$2.303 \log. \frac{P_2}{P_1} = \frac{L}{R} \left(\frac{1}{T_1} - \frac{1}{T_2} \right) \quad (1)$$

In this equation, P_1 represents the vapor pressure of the volatile liquid at the temperature T_1 , and P_2 at the temperature T_2 , where the temperatures are in absolute degrees ($F + 459.5$) and the pressures are in absolute units. Any units of pressure and temperature may be used providing the same units are adhered to. Thus the pressure may be expressed in pounds per square inch absolute, and the temperature in degrees Fahrenheit absolute. L is the latent heat of vaporization of the volatile substance, expressed as B.t.u. per pound molecular weight when T is in degrees F. absolute, or in calories per gram molecular weight when T is expressed in C. degrees absolute. $R = 1.99$ in either case.

Example. Water, whose molecular weight is 18, has a vapor pressure of 14.7 lbs. per square inch at 212° F. Estimate its vapor pressure at 150° F.

Solution. L for water at 212° = 970 B.t.u. per lb. $\times 18 = 17,460$ B.t.u. per pound molecularweight.

Therefore, using equation (1),

$$2.303 \log. \frac{P_2}{14.7} = \frac{17,460}{1.99} \left(\frac{1}{212 + 459.5} - \frac{1}{150 + 459.5} \right)$$

$$\text{Log. } \frac{P_2}{14.7} = -.579.$$

$$\frac{P_2}{14.7} = 0.264.$$

$$P_2 = 3.88.$$

The true vapor pressure of water at 150° F. from the steam tables is 3.71 lbs. per square inch. It is therefore evident that while this equation is valuable, it is only approximate and should not be used where steam tables or their equivalent are available. The equation is especially valuable for estimating the vapor pressure of liquids or solutions for which there are no accurate data, but should be used over short temperature ranges only.

* See W. Nernst: "Theoretical Chemistry," 4th Ed., 1916. Pp. 57 et seq.

The vapor pressure temperature relations for the most common pure liquids are known and may be found in such physico-chemical tables as those of Landolt-Börnstein, etc. In evaporation work, however, water is almost always the volatile liquid, and the engineer is most interested in the determination of the vapor pressures of the aqueous solutions under varying conditions of temperature and concentrations.

Raoult's Law. The vapor pressure of water, or of any other volatile liquid, is lowered by dissolving in it salts or other substances. The amount of the lowering is roughly proportional to the amount of the dissolved material present. This relation is given more precisely by Raoult's law, according to which the vapor pressure lowering is directly proportional to the mol fraction of the dissolved substance. Thus, in a solution containing 100 molecules or molecular weights, of which 95 are water and 5 are sugar, the mol fraction of the sugar is 0.05 and that of the water is 0.95. If n represents the mol fraction of the desired substance, P the vapor pressure of pure water at the temperature in question, and p is the vapor pressure of the solution at the same temperature, Raoult's law gives the equation

$$\frac{P - p}{P} = n \quad (2)$$

Example. What is the vapor pressure of a 40 per cent. sucrose solution at 212° F. according to Raoult's law, if the molecular weights of sucrose and water are 342 and 18 respectively?

Solution. The mol fraction of the sucrose is calculated as follows:

$$n = \frac{\frac{40}{342}}{\frac{40}{342} + \frac{60}{18}} = 0.0339.$$

The vapor pressure of water at 212° F. is 14.7 lbs. per square inch, therefore,

$$0.0339 = \frac{14.70 - p}{14.70}$$

$p = 14.20$ lbs. per square inch, which is the pressure desired.

The vapor pressure at any other temperature may be calculated in the same way. For instance, the vapor pressure lowering at 100° F. for the same solution as above, would be calculated as follows:

At 100° F., water has a vapor pressure of 0.942 lbs. per square inch. The vapor pressure lowering for the 40 per cent. sucrose solution would therefore be

$$0.942 - p = 0.0339 \times 0.942 = 0.032 \text{ lbs. per square inch.}$$

This method for estimating vapor pressures is quite accurate for organic solids such as sucrose dissolved in water when the mol fraction

is not greatly in excess of 0.05, but fails completely for inorganic salts, acids and bases, which are dissociated in aqueous solutions, and which give abnormally high vapor pressure lowerings. The only safe way to obtain the vapor pressure of such solutions, if not already available, is by actual experiment.

Boiling Point Raising. The reduction in the vapor pressure of a solution by the addition of dissolved solids requires that the temperature of the solution be raised in order to increase the vapor pressure to that of the pure liquid at the lower temperature. For instance, in the preceding example, the 40 per cent. sucrose solution has a vapor pressure of 14.20 lbs. per square inch at 212° F. If the external pressure were at one atmosphere, 14.7 lbs. per square inch absolute, at which pressure water boils at 212° F., the sucrose solution would not boil, but its temperature would have to be raised enough above 212° F. to raise the vapor pressure of the solution from 14.20 to 14.70 in order to cause the ebullition to begin. This increasing temperature is called the boiling point raising of the 40 per cent. sucrose solution at one atmosphere pressure, and is one of the factors in which evaporator engineers are vitally interested.

Calculation of Boiling Point Raising. The simplest expression for the boiling point raising at one atmosphere pressure for dilute solutions of the above type which follow Raoult's law approximately, is as follows:

$$\text{B.P.R. (Boiling point raising)} = 51.5 \, n \quad (3)$$

where n is the mol fraction of the dissolved substance as defined above, and 51.5 is the "boiling point constant" for water at 212° F.

Example. At one atmosphere pressure, the boiling point raising for the 40 per cent. sucrose solution would be

$$\text{B.P.R.} = 51.5 \times 0.0339 = 1.75^\circ \text{ F., approximately,}$$

and its boiling point would be

$$212 + 1.75 = 213.75^\circ \text{ F.}$$

The boiling point raising for other pressures may be calculated in the same way if the boiling point constant for the corresponding boiling temperature for pure water be known. This constant may be approximated in the following manner:

$$\text{B.P. Constant at temperature } t \text{ (F.)} = 51.5 \left(\frac{t + 459.5}{671.5} \right)^2 \quad (4)$$

where t is not far from 212° F.

Example. The B.P.R. for the 40 per cent. sucrose solution at a pressure of 0.942 lb. per sq. in. absolute, at which pressure water boils at 100° F., would be

$$\text{B.P.R.} = 51.5 \left(\frac{100 + 459.5}{671.5} \right)^2 \times 0.0339 = 1.21^\circ \text{ F.}$$

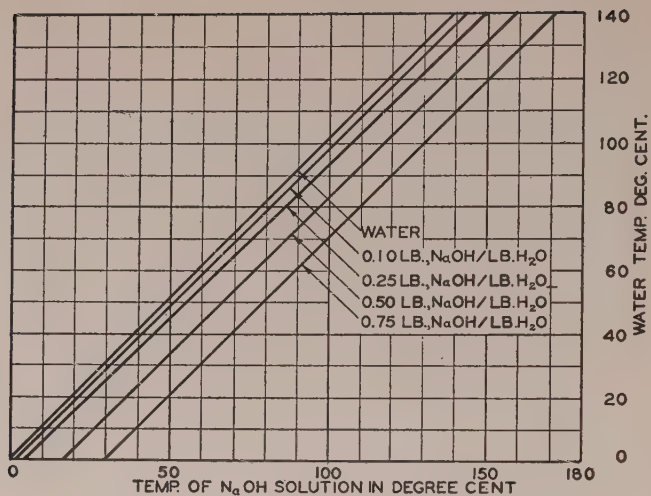


FIG. 1.

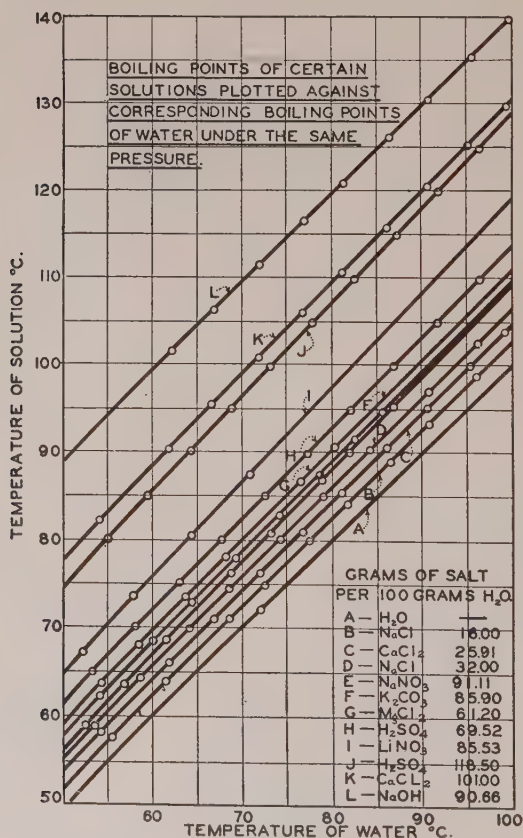


FIG. 2.

giving a boiling temperature for the solution at that pressure of 101.21° F., approximately.

From the above two examples it is evident that the boiling point raising for the same solution at two different pressures is not the same, and the usual assumption made by evaporator engineers, that it is a constant, is therefore a very rough one, frequently being in error by as much as 50 per cent.

Dühring's Rule. A graphical method for the estimation of boiling points of solutions is that illustrated in Fig. 1. (From Walker, Lewis and McAdams' Chemical Engineering.)

In this figure, each line shows the boiling temperatures of a specific caustic soda solution and the corresponding boiling temperature of water at the same pressure, plotted in Centigrade degrees. Thus, a solution containing 0.75 lb. of caustic soda per pound of water, or 43 per cent. solution, would boil at 130° C., when water boiled at 100° C., or at atmospheric pressure.

Similar charts may be constructed from data obtained from other solutions.

Fig. 2 (Trans. Am. Inst. of Chem. Eng. XIII, part 2, page 230, 1922) shows similar curves for a number of other aqueous solutions.

Since these boiling point curves are nearly straight lines, if two points on the boiling point curve be known, a straight line connecting them on a Dühring's rule diagram, will usually give an excellent approximation of the boiling points for intermediate pressures.

Chapter 3.

Transmission of Heat.

Evaporation can take place only by the expenditure of heat. Every pound of water that evaporates at 212° F. requires approximately 970 B.t.u.'s. (One British thermal unit, B.t.u., is the amount of heat needed to raise one pound of water at room temperature one degree Fahrenheit.) This heat can come from two sources only, from the vaporizing liquid itself, or from some external source, or from a combination of the two. If it comes from some external source, it must flow from that source to the vaporizing liquid in order to be used. It is the rate at which this heat flow takes place in which the evaporator engineer is interested.

There are three methods by which heat can be transferred from one body to another or from one part of a mass to another part of the same mass. One of these is by *conduction* where the heat passes through the matter without any portion of the matter itself being displaced. The second mode only applies to liquids and gases and is called *convection*. This refers to the transfer of heat by currents of liquid or gases moving from the cooler portion of the mass to the hotter, or conversely. The third method is *radiation* which refers to the propagation of heat by means of ether waves which is the manner in which we receive the heat of the sun.

Heat flows from one point to another because of a difference in temperature between the two points, and for no other reason. The greater the difference in temperature, the greater the rate of flow. Heat may flow by convection, in which case, the rate is directly proportional to the temperature difference. It may also flow by radiation in which event the rate of flow is proportional to the fourth power of the difference between the absolute temperatures. The evaporator engineer rarely has to do with the latter.

Heat flow by conduction is very like flow of electricity. In both cases, if the area of the cross section of the conductor be doubled, the amount of heat or of electricity which will flow through it will be doubled. If the resistance to flow be doubled, the amount of heat will be cut in half. It is therefore possible to make use of a simple equation to express the heat flow,

$$\frac{Q}{\Theta} = \frac{A\Delta t}{R} \quad (1)$$

Where Q represents the amount of heat flowing, Θ represents the time during which the flow continues, A is the cross sectional area of the con-

ductor, Δt is the temperature difference between the two sides of the heating surface, and R is its resistance. This equation is strictly true only for certain special conditions, but for all practical purposes is satisfactory for evaporation work if average values of the quantities involved be used.

It is not customary to make use of the resistance R in heat flow, although to do so simplifies computations in many cases. The reciprocal of the resistance H , the conductivity, is the quantity almost always used, giving the equation,

$$\frac{Q}{\Theta} = H A \Delta t \quad (2)$$

where average values of the quantities involved should be used as before. The proper averages will be discussed elsewhere.

Where heat flows through a flat conducting wall of uniform material with a constant difference in temperature between the two sides, the heat flow through any section of the wall will be the same as that through any other like section. If the wall be 1 foot thick, the amount of heat passing through each square foot of the wall per unit of time, the hour, will be half as much as if the wall were half as thick. It is therefore usual to write the heat flow equation in the form,

$$\frac{Q}{\Theta} = \frac{k A \Delta t}{l} \quad (3)$$

where l is the thickness of the wall in feet, and k is the "specific conductivity" of the material out of which the wall is made, expressed in the English system, as B.t.u.'s per square foot of cross section of the wall for a wall one foot thick, where the difference in temperature between the two faces of the wall is 1° F. per hour. Thus if a certain flat wall was constructed of material that had a specific conductivity of 100, and was 10 feet square and 2 inches thick, with a temperature difference of 50° F., between the two surfaces, the heat passing through it per hour, neglecting the edges, would be,

$$\frac{Q}{\Theta} = \frac{100 \times (10)^2 \times 50}{2/12} = 3,000,000 \text{ B.t.u.'s.} \quad (4)$$

If by chance the wall had not been flat but in the form of a cylinder as is the case of a pipe or tube, where the inside area per foot of length of pipe is less than the outside area, it is necessary to determine the average area of cross section of the wall per foot of length of cylinder in order to use it in the equation. For a cylindrical tube made of uniform material, the true average diameter is given by the equation,

$$D_{\text{average}} = \frac{D_o - D_i}{2.3 \log. \frac{D_o}{D_i}} \quad (5)$$

where D_o is the outside diameter, and D_i , the inside diameter of the wall of the tube. For thin wall tubes, it is permissible to use the arithmetic mean instead of the logarithmic mean,

$$D_{\text{average}} = \frac{D_o + D_i}{2} \quad (6)$$

Example. The specific conductivity of lead is 20. Calculate the amount of heat that would flow through a lead pipe 0.5 inch inside diameter with walls 0.5 in. thick, the pipe being 10 ft. long and the temperatures of the inside and outside surfaces being 250° F., and 100° F., respectively.

Solution. This problem consists largely in the determination of the average area of the wall of the pipe. The average diameter would be,

$$D_{\text{average}} = \frac{1.5 - 0.5}{2.3 \log \frac{1.5}{0.5}} = 0.911 \text{ ins.} \quad (7)$$

The average area would therefore be

$$\frac{0.911 \times \pi \times 10}{12} = 2.38 \text{ sq. ft.}$$

and the heat flow would be,

$$\frac{Q}{\theta} = \frac{20 \times 2.38 \times (250 - 100)}{0.5/12} = 171,000 \text{ B.t.u. per hour.}$$

In a great many cases, the walls through which the heat passes are other than flat or cylindrical, and if they are large compared with their thickness, it is usually satisfactory for purposes of evaporator design to

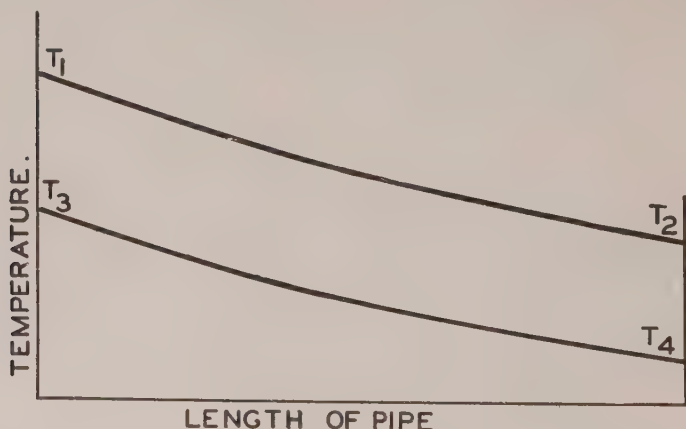


FIG. 1.

determine the arithmetic mean area and to use that in the equation of heat flow.

A variation in temperature in different parts of the heating surface requires the use of the average value of the temperature difference. The proper average to use depends upon the conditions. There are a number of possible cases, some of which are illustrated diagrammatically in Fig. 1 above.

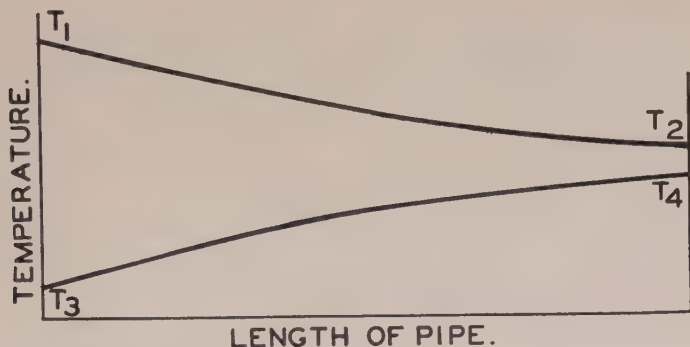


FIG. 2.

This diagram represents the length of a pipe where heat is passing through the wall either in or out as the case may be. The upper curve represents the change in the temperature of one wall of the pipe as it goes from one end to the other, while the lower curve represents the corresponding temperature of the other side of the pipe wall, the distance between the two curves giving the temperature difference between the two walls at any point along the pipe. In the case illustrated, the temperature difference at one end is $t_1 - t_3$ and the temperature difference at the other end is $t_2 - t_4$. If the temperature curves are smooth, as shown, or as in Fig. 2, the true average temperature difference is the logarithmic,

$$\Delta t_{\text{average}} = \frac{(t_1 - t_3) - (t_2 - t_4)}{2.3 \log \frac{(t_1 - t_3)}{(t_2 - t_4)}} \quad (8)$$

If, however, $t_1 - t_3$ is less than double $t_2 - t_4$, the arithmetic mean

$$\frac{(t_1 - t_3) + (t_2 - t_4)}{2} \quad (9)$$

may be used, with negligible error.

If the temperature curves are not smooth, as shown above, but if either one or both are broken as shown in Fig. 3, it is not correct to use average values of the temperature difference for the whole of the heating surface, but the heating surface must be considered in parts, such that

the temperature curves for each part are smooth. In the case illustrated, it is necessary to divide the length into two parts by the dotted line, and to calculate the two parts separately. A familiar type of the case is the condenser where cooling of the condensate occurs after condensation. In order to calculate the heating surface or the capacity, it is necessary to divide the heating surface into a condensing portion, and a cooling portion.

Applications of the use of these average values of area and of temperature difference will be made elsewhere, notably in the chapter on heaters. It should, however, be mentioned at this point, that in order

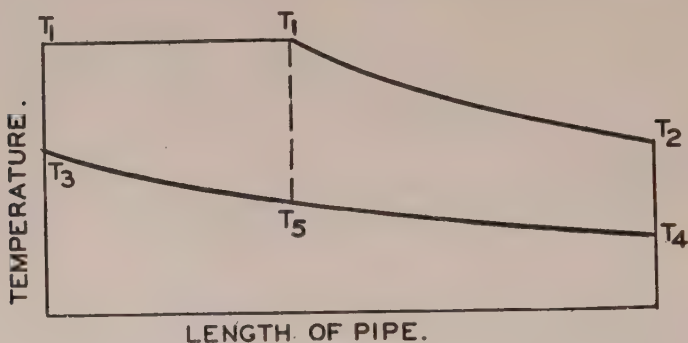


FIG. 3.

to use the above methods of calculating the amount of heat flow, it is necessary that the temperatures remain steady; that is, at any point in the apparatus, the temperature must remain unchanged. It is of course true that in any commercial apparatus, the temperature at any point will undergo fluctuations, and in such cases, it is necessary to determine the average temperature by taking the mean of a number of observations and using that mean value in the calculations.

So far the walls through which the heat is flowing have been considered to be made up of one uniform material only. It frequently happens, however, that the walls are composite, made up of several layers of materials of different specific conductivities. In such cases, it is convenient to think of resistances instead of conductivities, and to think of the total resistance of the wall as the sum of the resistances of the several layers. If the conductivity of a layer of uniform material is $\frac{k}{l}$, then the resistance of that layer is $\frac{l}{k}$ and the total resistance of the whole will be the sum of the resistances or

$$\frac{l_1}{k_1} + \frac{l_2}{k_2} + \frac{l_3}{k_3} + \text{and so forth,} \quad (10)$$

where the subscripts 1, 2, 3, etc., refer to the successive layers of the wall. The heat flow equation for such a case will therefore be

$$\frac{Q}{\Theta} = \frac{A\Delta t}{\frac{l_1}{k_1} + \frac{l_2}{k_2} + \frac{l_3}{k_3} \text{ etc.}} \quad (11)$$

The above equation applies to flat or nearly flat walls only. For cylindrical walls or walls of other shapes, it is necessary to determine the average area of each layer separately, and calculate the resistance of the whole of each layer, adding them all together afterwards. The heat flow equation then has the form

$$\frac{Q}{\Theta} = \frac{\Delta t}{\frac{l_1}{k_1 A_1} + \frac{l_2}{k_2 A_2} + \frac{l_3}{k_3 A_3} + \text{etc.}} \quad (12)$$

In nearly all cases, in evaporation, heat flows from one fluid through a separating wall into another fluid, the fluids being either liquids or vapors as the case may be. It has been shown that where such a fluid is in contact with a solid wall, there exists a film of the fluid immediately next to the wall, which is practically stationary and through which heat must diffuse by direct conduction.

The movement of heat through the main body of a liquid or vapor is usually rapid due to convection currents which convey the heat from one portion of the body of the fluid to another part by direct convection. But most fluids have low coefficients of heat transmission, and therefore, when the heat must flow through the fluid by conduction only, the rate is relatively low, and the surface film offers, therefore, a considerable resistance to the heat flow.

Where surface films exist, therefore, the total resistance to the flow of heat from one point to another must include the resistance of these films as well as those of the solid wall. The equation for heat flow becomes

$$\frac{Q}{\Theta} = \frac{\Delta t}{\frac{1}{h_1 A_1} + \frac{l_2}{k_2 A_2} \cdots \frac{1}{h_n A_n}} \quad (13)$$

where h_1 is the conductivity of the fluid film on one side of the wall, the area of the surface on that side being A_1 , and h_n is the conductivity of the fluid film on the other side of the wall where the area of the film is A_n , the intermediate terms in the denominator being the resistances of the several sections of the composite wall as in the previous equation. Illustrations of the use of these equations will be given in later chapters.

TABLE I.

(Table from Walker, Lewis and McAdams' Chemical Engineering.)

Specific conductivities of common substances,

COEFFICIENT OF HEAT CONDUCTIVITY OF SOLIDS.*

(k = B.t.u./hr./sq. ft./° F. difference°/ft. thickness.)

(a) Metals.

Material	Experimental Temperature Range ° F.	k	Reference Number
Aluminum	32-212	85	I
Brass, yellow	32-212	55	I
Copper, pure	at 68	238	I
Iron:			
Pure	at 76	37	I
Wrought	32-527	40	I
Cast, 3.5 per cent. C.	at 86	36	I
Lead	at 59	20	I
Nickel	68-392	32	I
Platinum	64-212	41	I
Silver	at 64	243	I
Steel	at 82	25 §	
Tin	at 32	35	I
Zinc	at 64	64	I

(b) Non-Metals.

Asbestos	100-1,000	0.12	2
Brick:			
Carborundum	at 1,800	3.4 †	
Magnesia	at 1,800	1.8	2
Fire	at 1,800	0.75 †	
Building	at 1,800	0.8	2
Silica	at 1,800	0.5	2
Kieselguhr	at 1,800	0.4	2
Sil-O-Cel	at 1,800	0.03	2
Brick and mortar wall.		0.4 †	
Portland cement, neat	at 95	0.5	2
Cork	122-392	0.03	2
Electrode carbon	212-1,700	32 ‡	2
Glass, flint	50-59	0.3	2
Infusorial earth (12.5 lbs. per cu. ft.)	at 122	0.05	2
Leather	72-97	0.1	I
Mineral wool	70-350	0.03	2
Magnesia	68-310	0.04	2
Rubber	at 220	0.1	I
Sand	68-310	0.2	2
Wood, pine:			
Parallel to fibers.		0.07	I
Perpendicular to fibers.		0.02	I

* The presence of impurities, especially in the metals, may cause variation in the coefficient of conductivity of almost 100 per cent.

† Average of results from several sources.

‡ The coefficient of conductivity of this material is given as only 10 per cent of this value by other experimenters.

§ This is the lowest value usually quoted.

¶ "Landolt-Börnstein Tabellen," 1905 Ed.

|| CARL HERING, *Met. and Chem. Eng.*, 9, 652 (1911).

* "Landolt-Börnstein Tabellen," 1905 Ed.

† CARL HERING, *Met. and Chem. Eng.*, 9, 652 (1911).

‡ Smithsonian Tables, p. 217, 1920 Ed., Smithsonian Inst., Washington, D. C.

§ MARKS' "Mechanical Engineers' Handbook," p. 306, 1916 Ed., McGraw-Hill, New York.

(c) Gases.

	At 32° F.	At 212° F.	Reference Number
Air	0.0137	0.0174	3
Carbon monoxide	0.0131		3
Carbon dioxide	0.00804	0.0120	3
Oxygen	0.0138	0.0180	3
Nitrogen	0.0138	0.0174	3
Hydrogen	0.0101	0.0121	3
Methane	0.0174		3
Steam	0.0095	0.0129	4

TABLE II.

(Table from Walker, Lewis and McAdams' Chemical Engineering.)

THERMAL CONDUCTIVITY OF LIQUIDS.

Substance	Thermal Cond. (<i>k</i>)	Sp. Gr. (<i>s</i>)	Sp. Ht. (<i>cp</i>)	Mol. Wt. (<i>M</i>)	$\sqrt[3]{\frac{M}{s}}$	$\frac{k}{sc} = \alpha_1$	$\frac{k}{sc} \sqrt[3]{\frac{M}{s}} = \alpha_2$
Methanol	0.12	0.81	0.586	32	3.41	0.253	0.862
Ethyl alcohol	0.1025	0.804	0.579	46	3.86	0.220	0.850
Propyl alcohol	0.0902	0.819	0.518	60	4.18	0.213	0.890
Chloro-benzene	0.0731	1.111	0.352	112.5	4.65	0.187	0.870
Carbon tetrachloride .	0.0610	1.63	0.205	153.8	4.56	0.182	0.833
Benzene	0.0806	0.873	0.460	78.0	4.48	0.201	0.899
Aniline	0.0990	1.039	0.497	93	4.47	0.192	0.857
Toluene	0.0742	0.884	0.440	92	4.71	0.191	0.899
Carbon disulphide ...	0.0647	1.263	0.240	76.1	3.93	0.214	0.840
Chloroform	0.0697	1.503	0.234	119.4	4.30	0.198	0.852
Water	0.329	1.000	1.006	18	2.62	0.329	0.862
Glycerol	0.155	1.26	0.576	92	4.18	0.214	0.892
Acetic acid	0.1141	1.069	0.487	60	3.81	0.219	0.835
				Av.....			0.864

Chapter 4.

Conductivity of Surface Films.

It was pointed out in the preceding chapter that the flow of heat from one fluid to another through a solid dividing wall was influenced by the resistance to flow of heat of fluid films on the surface of the solid wall. The resistances of these films has been the subject of a great deal of study and experiment, notably by W. H. McAdams at the Massachusetts Institute of Technology, most of the following material having been obtained from him.

Non-Boiling Liquids. The latest work on the resistance of the film of liquid on a dividing wall where the liquid is flowing through a circular pipe under forced circulation and is not boiling indicates that the film conductivity may be expressed by the following equation:

$$h = \frac{5.06}{D^{0.2}} \left(\frac{V}{z_f} \right)^{0.8} \left(1 + \frac{50}{r} \right) R$$

In this equation, h is the conductivity of the film expressed in English units, B.t.u. per square foot of surface per degree Fahrenheit temperature difference per hour. D is the inside diameter of the pipe in inches. V is the average mass velocity of the liquid flowing through the pipe, expressed as pounds per square foot of cross section of the pipe per second. z_f is the relative viscosity of the liquid film, relative to water at 68° F. r is the ratio of the length of the pipe to its inside diameter, and R is the ratio of the thermal conductivity of the liquid to that of water (Water = 0.329 approximately). (See Table II.)

Example. Calculate the conductivity of the water film on the inside surface of a pipe 2 ins. in diameter and 20 ft. long, if the average relative viscosity of the water film on the surface of the pipe is 1.5, and it is flowing through the pipe at a linear velocity of 5 ft. per second.

Solution. First calculate the mass velocity, V , from the linear velocity, u .

$$V = \rho u$$

where ρ is the density of the liquid expressed as pounds per cubic foot, in this case about 62.4. Also $R = 1$

$$V = 62.4 \times 5 = 312.$$

$$h = \frac{5.06}{2^{0.2}} \left(\frac{312}{1.5} \right)^{0.8} \left(1 + \frac{50}{\frac{20 \times 12}{2}} \right)$$

$$h = 444.$$

It should be noted in connection with this and the other equations for film conductivities, that the relative viscosity of the film is determined by the temperature of the film itself, and not that of the average temperature of the liquid in the pipe. The determination of the temperature of the film is not easy, but it is usually true that it is possible to estimate this temperature very closely if the temperature of the liquid in the pipe and the wall temperature be known, by assuming that the film temperature is half way between the two. Where the wall or liquid temperatures change from one point to another as is usually true in heaters and coolers, it is necessary to determine the average temperature of the film, and this may for all practical purposes be taken as the arithmetic mean between the averages at the two ends of the apparatus.

The film conductivity for non-boiling liquids which are moving by the solid surface through which the heat is passing by convection currents only and are not being pumped in any way, depends to a considerable extent on the arrangement of the heating surface, the depth and size of tank, the viscosity of the fluid in the tank, and the difference in temperature between the heating walls and the solution. For small copper pipes in coils immersed in large tanks with steam inside the coils and aqueous solutions in the tank, there has been derived the following equation based on experiments.

$$h = 35 J^{0.8} \Delta t^{0.4}$$

This equation should give approximate results for such work, but should be used with a considerable factor of safety. The term J is the fluidity of the liquid which is being heated, based on the average temperature of the liquid in the tank and not on the film temperature as in the previous case. It is the reciprocal of the relative viscosity. The term Δt represents the difference in temperature between the surface of the hot wall and the average temperature of the liquid in the tank, in degrees F.

Example. Molasses with a viscosity relative to that of water at 68° F. of 100 is being heated in a large open tank by means of copper steam coils. The temperature of the molasses is 100° F. and that of the outside wall of the coil is 200° F. Estimate the conductivity of the molasses film on the outside surface of the coil.

Solution. The fluidity J is the reciprocal of the relative viscosity, therefore

$$J = \frac{1}{100}$$

$$h = 35 \left(\frac{1}{100} \right)^{0.8} (200 - 100)^{0.4} = 55, \text{ ans.}$$

Condensing Vapors. The film conductivity of condensing vapors is generally very high. Values are given in the literature for steam under favorable circumstances of 4000 in English units, and it is believed that higher values may be obtained.

The mechanism by which vapor condenses when coming into contact

with a cold wall may be understood when it is realized that condensing vapor loses practically all of its volume in changing from the vapor to the liquid state. This sudden decrease in volume causes an instantaneous rush of additional vapor to take its place with the result that the velocity with which the steam is brought up to the condensing surface is of relatively less importance than is the case with non-boiling liquids where the steam is pure containing no non-condensable gases. Where such gases are present, as is frequently the case, a higher vapor velocity is of great importance as will be pointed out later.

Once the vapor has condensed and has formed a film of liquid on the condensing surface, further flow of heat from the vapor wall must take place through this film and therefore the thermal conductivity of the liquid is of importance. Furthermore, the liquid must run off from the surface as rapidly as it forms in order to avoid an accumulation and resulting insulation of the surface with a thick layer of condensed vapor. This means that the fluidity of the liquid at the temperature at which it exists on the surface is also of great importance. It has been shown by McAdams that for pure vapors containing no non-condensable gases, the film conductivity may be expressed with a considerable degree of precision by the equation,

$$h = 2200 KJ.$$

where h is the film conductivity of the condensing vapor in the usual English units, K is the thermal conductivity of the condensed vapor in English units, and J is the fluidity of the condensed vapor at the temperature at which it exists on the condensing surface.

Example. The film conductivity of condensing steam at 212° F. is calculated as follows:

K for water may be taken to be 0.329

J for water at 212° F. is 3.6

$$h = 2200 \times 0.329 \times 3.6 = 2600.$$

The film conductivity of steam condensing under a 28-inch vacuum where the fluidity of water (at 101° F.) is approximately 1.5, is $2200 \times 0.329 \times 1.5 = 1100$.

The effect of lowered temperature on film coefficients is well illustrated by the above cases, and is of the greatest importance in evaporator work. The evaporator engineer rarely has to do with liquids other than aqueous solutions, but when such a case occurs, the proper values for K and J may usually be found in physical-chemical tables such as Landolt-Börnstein, etc.

Boiling Liquids. The situation with respect to the film coefficient for boiling liquids is very unsatisfactory at the present writing, 1925. Heat transmission has been the subject of extended research for a great many years and accurate tests on evaporators are available as far back as 1890. But in nearly all cases, until recently, overall coefficients of heat transfer, that is, from the condensing steam to the boiling liquid in the case of

evaporators, have been determined and not the individual film resistances or coefficients. This is particularly unfortunate since it means that the coefficients so determined are of value only for the particular apparatus and solution tested and it is almost impossible to make use of them for other conditions. But on the other hand, from the standpoint of actual design, the situation is not as serious as the above statements would make it appear for the following reasons:

Commercially, solutions which are subject to evaporation in nearly all cases contain substances, either dissolved or in suspension, will give

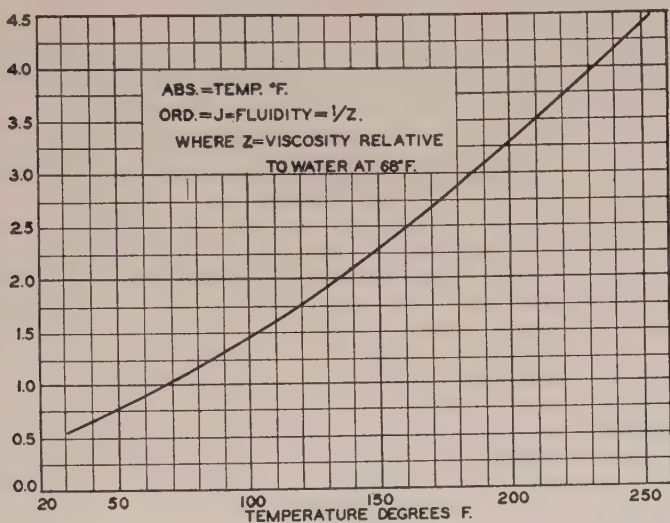


FIG. 1.

deposits on the heating surfaces. These scales are frequently of such a character that the overall coefficient of heat transfer and the capacity of the apparatus will decrease to from one-quarter to one-tenth of its clean rating in one week's operation or less. Under these circumstances, the best estimate of coefficient of heat transfer when clean is of considerably less importance than an accurate knowledge of how rapidly the surface is going to foul. This information can come only from actual experience in handling the material in question and any design for a material of which the fouling characteristics are not well known must include a large factor of safety in order to insure full capacity under worst possible conditions. An estimate of the rate of scale formation will be discussed later.

All that can be said of the film coefficient for boiling liquids at the present time is purely qualitative. Experiments which have been carried on at the Massachusetts Institute of Technology indicate that the film conductivity is affected by several factors. These include the fluidity of

the boiling liquid, the temperature difference between the boiling liquid and the surface of the hot wall, the thermal conductivity of the boiling liquid and its latent heat of vaporization, and the velocity of the liquid by the heating surface. It is hoped that within a short time, experiments which are now in operation will furnish sufficient material so that boiling film coefficients may be predicted with the same degree of assurance that is possible with non-boiling liquids. In the meantime, it is necessary to work with overall coefficients, using experience and good engineering judgment in the interpretation of overall values previously obtained by others.

Superheated Vapors and Gases. The film conductivity for superheated vapors and gases is very low. Values of the order of magnitude of from 5 to 10 are usual for gases under moderate pressures and velocities, and superheated vapors are supposed to have similar values so far as is known at the present time. In evaporator practice where steam is used almost exclusively for the heating medium in everything except the crudest of fire heated evaporators, gases, as such, play a small part. Superheated vapors, however, are of frequent occurrence due to the fact that the vapors liberated from the surface of a boiling liquid which has in it dissolved substances which have raised its boiling point above that corresponding to that of the pure liquid at the same pressure, are superheated to the extent of the boiling point raising. These more or less superheated vapors, when taken to the steam chest of the following effect or to the surface condenser, have the low value of the film coefficient noted above until they have lost their superheat, after which they have the larger values of saturated vapors.

Evaporator engineers have found that if the vapor so superheated be considered as saturated vapor at the lower temperature corresponding to its pressure and the corresponding temperature difference used with the coefficient for saturated vapor, the area of heating surface so calculated usually is found to be sufficient. This method is dangerous to use however where the degree of superheat is high, and the safer thing to do under such circumstances is to consider the loss of the superheat as a separate operation from the condensation of the saturated vapor, and to figure the two heating surfaces separately, adding the two together. It is believed that this method will introduce a reasonable factor of safety.

Non-Condensable Gases. Scale and non-condensable gases are the most troublesome factors with which the evaporator specialist has to deal. Steam containing air or other gases which do not condense at the temperature of the cooling surface has a materially smaller film conductivity than the pure vapor under the same conditions. This is due to the fact that the gas tends to adhere to the condensing surface and to insulate it from the condensing vapor. Since gas is a poor conductor of heat, an appreciable amount of it may render the cooling surface practically inoperative unless it is removed as fast as it collects.

The results of tests by Professor Kerr, *Trans. A.S.M.E.* 35, (1913), 731, on the effect of air in steam has been recalculated by the authors to a film basis and the following equation derived:

$$\text{Log}_{10} h = 1 + 0.0246X.$$

Where h is the film coefficient of the condensing steam and X is the average per cent. steam in the gas by volume during the condensing process as determined by measuring the temperatures and pressures in a number of points in the steam chest of the evaporator and taking the average of them, the data including average percentages from 100 to 60.

Example. In a certain evaporator, steam containing 1 per cent. air by volume is condensed in the steam chest, the uncondensable gases which are removed from the farther end of the steam chest having a steam content of 70 per cent. by volume. Calculate the film coefficient of heat transmission for the condensing steam.

Solution. The average steam content of the mixture in this case is a complicated function of a number of variables, but where the difference between the percentages of steam at the two ends is not too great, the logarithmic mean percentage may be used.

$$X_{\text{average}} = \frac{99 - 70}{\log_e \frac{99}{70}} = 83.$$

where $\log_e = \log_{10} \times 2.303$

$$\begin{aligned} \log_{10} h &= 1 + 0.0246 \times 83 = 3.04. \\ h &= 1100. \end{aligned}$$

The above equation is very approximate and should be used with that fact in mind until more exact information is available. A number of experiments at the Massachusetts Institute of Technology have indicated, however, that it is probably near the truth.

Chapter 5.

Factors Affecting the Coefficient of Heat Transmission.

In evaporators, the determination of the amount of heating surface to accomplish a given task under prescribed temperature conditions is most difficult, and resolves itself into the establishment of the proper coefficient of heat transmission, or the amount of heat that will flow through a square foot of heating surface in B.t.u.'s per hour per degree F. difference between the temperature of steam or vapor on one side and the temperature of the boiling liquid on the other.

The range of coefficients observed in practice, and noted in research work conducted by various investigators varies between very wide margins, from almost zero to as high as 2000, and it becomes a vital matter to consider carefully the reasons for these variations.

It is known that many factors influence this coefficient of heat transfer as determined by the total amount of heat transmitted per hour, the temperature of steam on one side of the heating surface, and the temperature of the liquid on the other side, represented by the reading of a thermometer inserted at the boiling level in the evaporator. Among these factors may be mentioned the following: 1) cleanliness of heating surface; 2) temperature of steam; 3) rate of evaporation; 4) hydrostatic head; 5) viscosity of the solution being concentrated; 6) presence of non-condensable gas; 7) thickness and material making up the heating surface through which heat must pass.

Some of these variables are so intimately related that it is impossible to separate them and thus isolate their influence for individual study and consideration. In practice, it is necessary to follow precedent, rather than be guided by considerations based on incomplete theory, which are very often misleading. For this reason, in the general discussion of applications to various industries approximate safe ratings have been given based on what is believed conservative data for the work outlined.

1) Cleanliness of Surface. In the discussion on the subject of Foul-ing of Heating Surfaces, it has been shown how largely heat transmission can be affected by the accumulation of scale and dirt on the liquor side of the tubes. Illustrations were given of tests wherein, with scale forming solutions, the depreciation of the coefficient of heat transmission was as great as 50 per cent. in a period of only two hours, and very much more for operating periods of the apparatus from normal clean out to normal clean out. Thus it is very apparent that this consideration alone can com-

pletely vitiate deductions arrived at on a small scale with ideal conditions where artificial solutions are used free from contamination usually encountered in practice. The reader must therefore be very wary of so-called high efficiencies and miraculous performances based upon tests carefully controlled and conducted under ideal conditions never found in the field.

It has been suggested in other pages that the safest way to guard against foolish optimism, where practice has not been thoroughly established, is to study the problem first hand with a pilot plant, wherein the actual con-

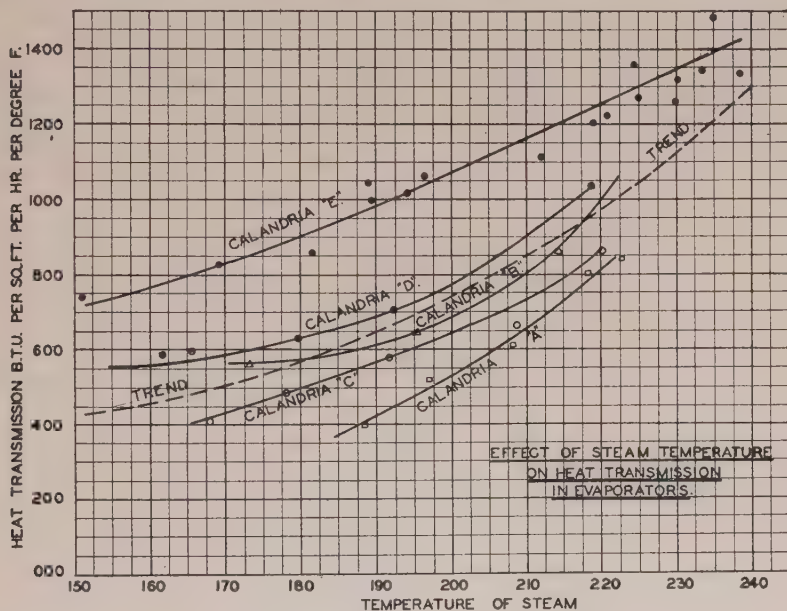


FIG. 1.

ditions of operation are faithfully reproduced, with all of the complications, and to so carry on the work that the facts are thoroughly established by repeated operations extended through sufficiently long periods of time, without minimizing or disguising the difficulties encountered.

2) **Temperature of Steam.** At first thought, it might be expected that a given temperature difference between the two sides of the heating surfaces of an evaporator should give the same transmission of heat, irrespective of the level at which this difference is taken. Theoretically, this is not so, and actually, it has been well established by many tests that higher transmissions are obtained at higher temperatures. E. W. Kerr carried on many experiments to establish this fact, and Fig. 1 gives the result of some of his research work in this direction. A study of these

curves shows that at a steam temperature of 230°F ., the heat transmissions are about double those shown at 150°F .

Researches on evaporation by H. K. Moore have indicated practically the same thing. Caution should be used in accepting these data. It was pointed out before that the variables are hard to separate. In making this range of tests, which were conducted with an operating difference of

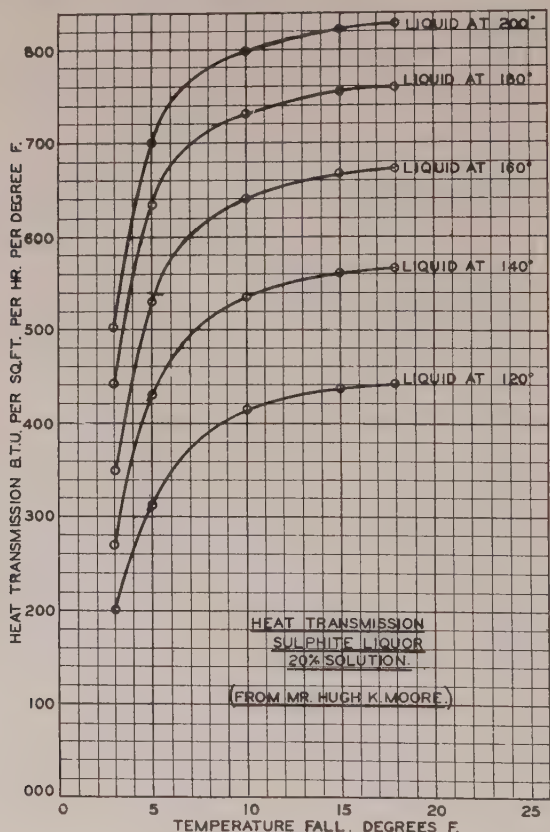


FIG. 2.

about 20°F ., the temperature on the liquor side of the evaporator varied considerably, from 130°F . to 210°F ., and no less than three other factors varied at the same time, all affecting the coefficient of heat transmission. First, the rate of evaporation changed, altering the transmission. Second, the hydrostatic head was constant, but affected the temperature difference, and hence the gross heat transmission, more adversely at low than at high temperatures, inasmuch as equal differences of pressures correspond to greater differences of temperatures at the lower range. Third, the viscosity

of the water being evaporated changed, making higher transmission at higher temperatures. It is almost impossible to segregate these influences and make corrections for them, therefore, the data brought out should be used merely as a guide, bearing in mind the facts as brought out above.

3) Rate of Evaporation. It has also been established experimentally, and confirmed in practice, although the data are not always consistent, that as the rate of evaporation is increased, the coefficient of heat transmission increases also. This may be due to several causes. In the first place, an increase of rate of evaporation in a given apparatus operating at a given temperature on its liquor side can only be obtained by raising the temperature of steam, and we have seen that raising this temperature increases the coefficient of heat transmission, although the effect in this case should not be very great. There is to be found a partial explanation in the higher velocities of travel in the tubes corresponding to the increased rating, as it has been shown beyond doubt that heat transmission increases as the circulation is increased. (See below and also chapters on circulation and solution heaters.)

Fig. 2 gives a graph made up from data taken from H. K. Moore's paper. In full-sized standard evaporators with circulation controls, an increase in the rate of evaporation would automatically be accompanied with a reduction of the static head, which constitutes another disturbing factor of importance. No doubt, a well-operated hand-controlled evaporator would follow the same variations, as the operator works on froth levels as a rule, and not by the water gauge glass indicating the levels of liquor in the evaporator.

For apparatus of the Lillie type, there would be no alteration of the static head, which is practically negligible in all cases. In the case of the Moore evaporator, there is no static head, but there is a back pressure corresponding to the dynamic head, which is substantially the same thing, and this would be increased with higher ratings, on account of higher speeds. (See chapter on Natural Circulation.)

4) Static Head. During the last ten or fifteen years, there has been quite a volume of written and oral discussion as to the influence of the depth of liquor upon the performance of evaporators. It has been established beyond any doubt that excessively high levels of liquor in evaporators have a very decided adverse effect upon the heat transmission. Here again, Professor Kerr has given us some very valuable information, and the plat herewith shows the results of experiments conducted by him, throwing light upon the subject. The data establish the fact that the level of the liquor in evaporator tubes should be maintained just high enough so that under conditions of evaporation, there is sufficient liquid thrown above the tops of the tubes to keep the surfaces thoroughly wetted at all times, and in addition, provide enough liquor circulation to avoid local concentration, as well as provide sufficient liquor circulation at the bottom of the tubes to stop these from fouling excessively. (See chapter on Natural Circulation.) The loss imposed by excessive liquor depth is greater even than it should be on account of the rise in the boiling point due to the pressure thus superposed upon that existing in the vapor space.

It is believed, however, that the problem has been greatly exaggerated, and private opinions have been set forth as facts, which, in view of actual operation in practice, have very scant foundation. For instance, it has been stated that the right level is one third the height of the tubes, and from this the conclusion has been assumed that it is impossible to get high heat transmissions from long tube evaporators. The facts are that for

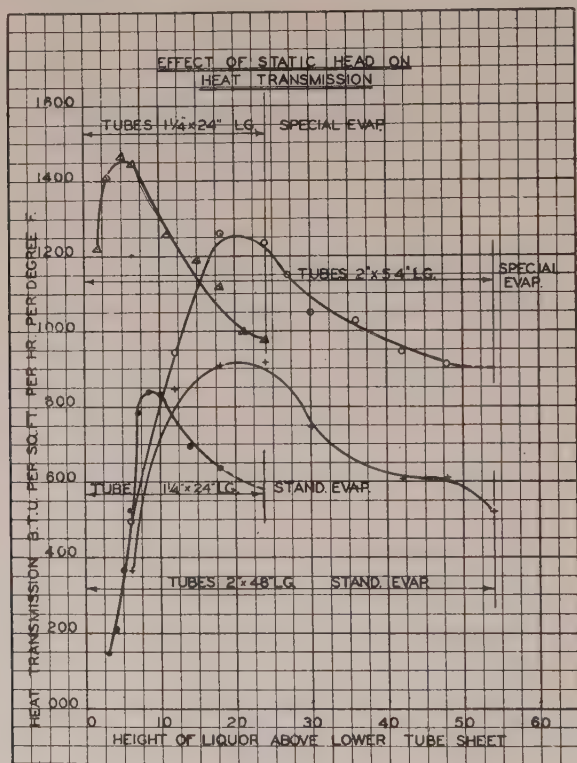


FIG. 3.

certain short tube evaporators, operating under certain conditions, with certain solutions, it has been found that carrying the levels one-third the height of the tubes gives good results, but no further conclusions are justifiable without the facts to support them.

Published tests of Kestner evaporators having very long tubes have shown that heat transmissions as high as 1147 B.t.u.'s per square feet per hour per degree F. drop, under commercial operating conditions, an exceedingly high rating even if the machine was clean when the test was made. A long tube apparatus designed by one of the authors has recorded heat transmissions of 1200 operating at 212, with only 9° F. difference

between steam and liquor, with the level 10 per cent. of the height of the tubes. In fact, where the incoming liquor is at or above the boiling point, for this type of apparatus, it is probable that there would be considerable loss of capacity by exceeding a level of 10 per cent. of the length of the tubes.

The Lillie evaporator is about the only one in which the static head has been practically completely eliminated. In spite of this, the performance of these evaporators as reported in published tests by reputable men do not show unusually high heat transmissions, especially when it is remembered that in almost every case, the operations were conducted with the surfaces perfectly clean, and therefore under most favorable conditions, which, as is well known, does not truly indicate the useful dependable commercial capacity of the evaporator.

In every case, the elimination of the static head has been attempted by introducing a mechanical pumping of the solution over the heating surface, for which there is necessarily a debit charge, both in cost of operation, and complication.

Therefore, one should not be too hasty in deviating from established practice, nor too willing to give up standards established after many years of experiment. As has been pointed out in another part of this book, the static head can be maintained at very reasonable proportions by controlling the circulation instead of the level.

5) Viscosity. It is to be regretted that our present information on the isolated influence of viscosity upon heat transmission in evaporators is very meagre. This is due to the fact, as was stated before, that it is extremely difficult to determine accurately on account of the ever present other variables. Experiments on heat transmission in heaters have shown that viscosity has a great influence on the coefficient of heat transmission. In evaporators, the effect is probably similar, but not exactly the same. Some information is available in H. K. Moore's papers, a plat of which is given in Fig. 4. There is a wide gap between viscosity 1.85, and viscosity 9.5 in which there are no readings. The general trend of the curve is plain enough, however, and is very illuminating. More elaborate data will be available later when the results of research work now being carried on are published.

6) Non-Condensable Gases. The ill effect of non-condensable gases is pointed out in another chapter in this book. (See chapter on Non-Condensable Gas Removal.) Besides interfering with heat conductivity, usually, there is erosion of the heating surfaces, as these gases are very often of a corrosive nature. The provisions against this difficulty are fully outlined in the above reference. With proper precautions, evaporators should not be affected adversely by non-condensable gases.

7) Thickness and Material of Surfaces. The matter of metal thickness in heating surfaces has been fully elaborated in other parts. For ordinary purposes, heat transmission is almost unaffected by either the composition or the thickness of the metal making up the heating surface, except in very special cases, such as tin-lined, or glass-lined apparatus, wherein the resistance to the free passage is so much increased that it

must be taken into consideration. These conditions are unusual, and the proper surfaces must be determined from experiments covering the actual proposed conditions.

Other Data. A very large amount of work on the transmission of heat in evaporators is available in the literature, notably by such investigators as Claassen, W. L. Badger, and many others. In almost every case, as has been pointed out previously, the information is in the form of overall coefficients and not as film coefficients. It is possible to analyze

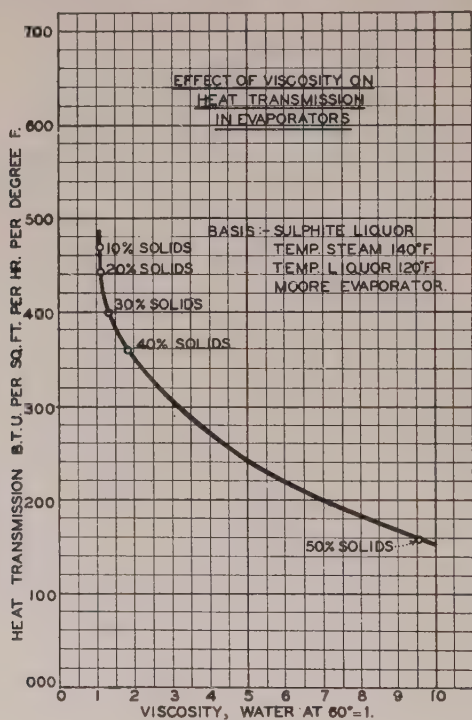


FIG. 4.

these results to a certain extent, however, due to the fact that in most commercial evaporators where viscous solutions are being handled, the bulk of the resistance to heat flow occurs through the film of boiling solution, the resistance of the condensing steam and that of metal of the tube being relatively small. Under such circumstances, it is possible to make a study of the effect of several variables on the overall coefficient. For instance, McAdams has taken certain such data and plotted on logarithmic paper the temperature difference between the steam and the boiling liquid as abscissa, and as ordinate, the product of the overall coefficient and the relative viscosity of the boiling liquid. For each set of tests on

the particular machine tested, the data gives a reasonably straight line. Fig. 5 is taken from "Principles of Chemical Engineering," Walker, Lewis and McAdams.

A number of interesting conclusions may be drawn from a diagram of this sort. For instance, for any given evaporator, the coefficient of heat transmission under a fixed temperature difference varies inversely as the relative viscosity of the boiling liquid.

If a certain type of evaporator when operating under a fixed temperature difference of 30° F. was found to have an overall coefficient of 750 when evaporating water at one atmosphere pressure where the relative viscosity of the water was 0.3, the same evaporator with the same tem-

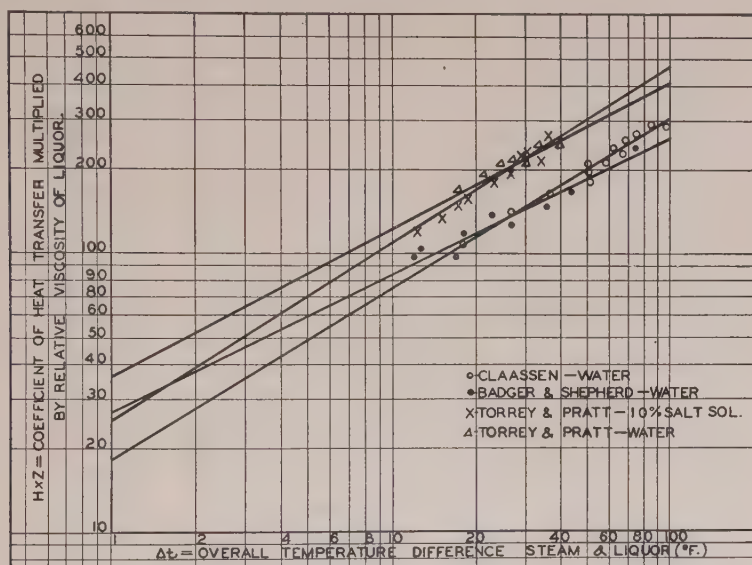


FIG. 5.

perature difference of 30° F. might be expected to have an overall coefficient of 225 when evaporating a 40 per cent. sucrose solution at 200° F. where its relative viscosity is approximately 1.0. Such an estimation presupposes satisfactory conditions on the condensing steam side with respect to non-condensable gases, and clean tubes. Another way of looking at it, and probably the more accurate, would be that the evaporator would have the same coefficient handling water with a temperature drop of 5° to 6° that it would have handling the sucrose solution with a 30-degree drop.

It has been pointed out that it is the property of such a plot on logarithmic paper, that, for instance, if, when the temperature difference is changed from 100 to 50, the HxZ product changes from 300 to 200. Halving

the temperature difference at any point will reduce the Hs product in the ratio of 300 to 200. The effect of this may be shown in the following example:

Suppose a triple effect evaporator is operating between certain temperature limits for its terminal conditions on some specific solution. Suppose again that it is proposed to double the number of effects with the same surface per effect concentrating the solution between the same limits with the same terminal temperatures as in the triple effect.

Doubling the number of effects would for all practical purposes halve the temperature drop for each body. The average viscosity of the solution

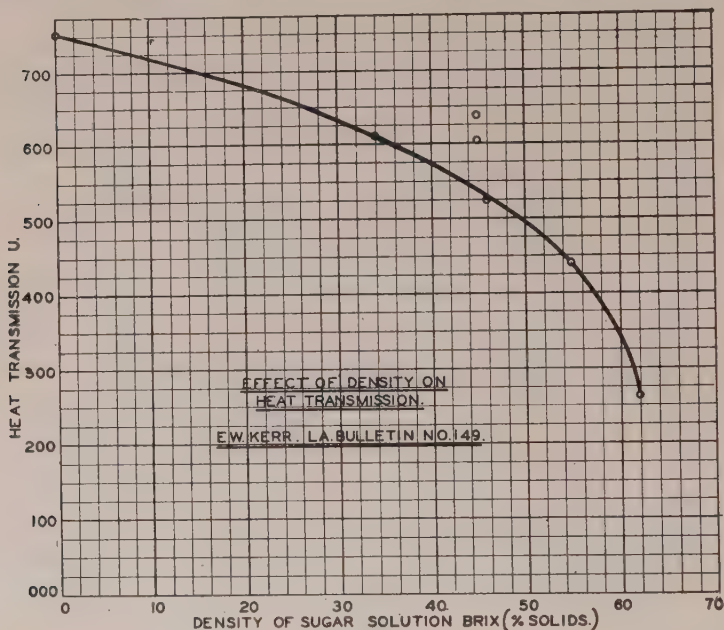


FIG. 6.

in the whole apparatus would not be changed appreciably. If the figures given in the earlier paragraph applied to the type of evaporator in question, then the coefficient of heat transfer might be expected to be reduced to two thirds of what it was before, and the overall capacity of the six effect evaporator would be $1/2 \times 6/3 \times 2/3 = 2/3$ of that of the triple. The steam consumption would, of course, be cut in half per pound of water evaporated, approximately.

The authors have no hesitation in asserting that it is at the present state of the art, impossible to predict accurately the coefficient of heat transmission for a given evaporator operating on a given solution under specific conditions unless previous experience or data are available. It

is well worth while, however, at this point to show some of the results of investigators of this important and complicated problem.

In *Bulletin 149* of the Louisiana State Agricultural Experiment Station, E. W. Kerr shows several diagrams of interest, some of which have been reproduced in the preceding discussion. Two more are given herewith in Figs. 6 and 7.

Fig. 6 shows the effect of the viscosity of a sugar solution on the coefficient of heat transmission. In this diagram, which is for a specific evaporator, operating under specific conditions, pure water was found to have a coefficient of about 750 B.t.u.'s per square foot per hour per 1° F. temperature difference, while with a 60 per cent. sugar solution, the coefficient had dropped down to about 250.

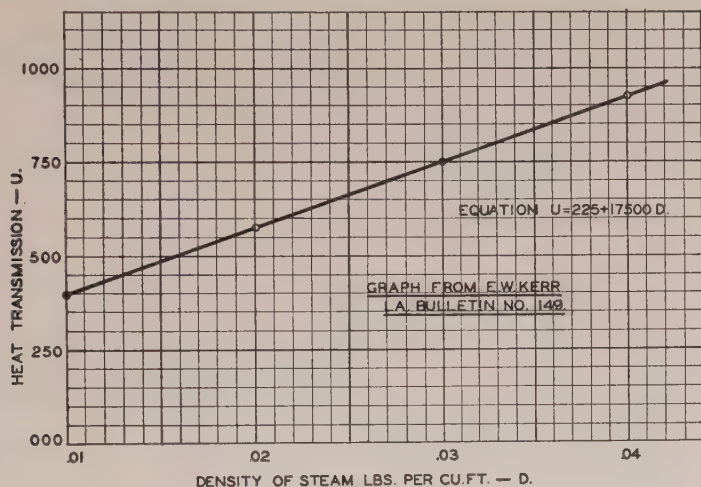


FIG. 7.

Fig. 7 shows the effect of increasing the steam pressure and temperature on the heat transmission for a specific case. This curve represents the average of a number of experiments on several evaporators, and therefore has no specific value except as indicating in a general way what might be expected on raising the temperature of steam. In this connection, it is felt that the method of attacking this particular problem given in the previous chapter, by the use of the film coefficients, cannot but prove more useful and more reliable.

Kerr, as contrasted with Moore, finds that in his experiments, the temperature difference had no effect on the overall coefficient of heat transmission. This is another example of the futility of attempting to predict the behavior of one type of evaporator when working on one solution, from the tests performed on another entirely different type of evaporator and another kind of solution.

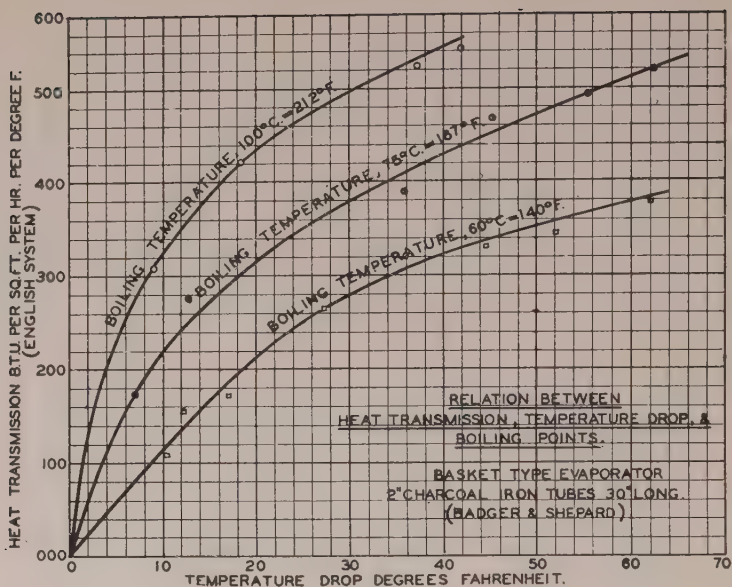


FIG. 8.

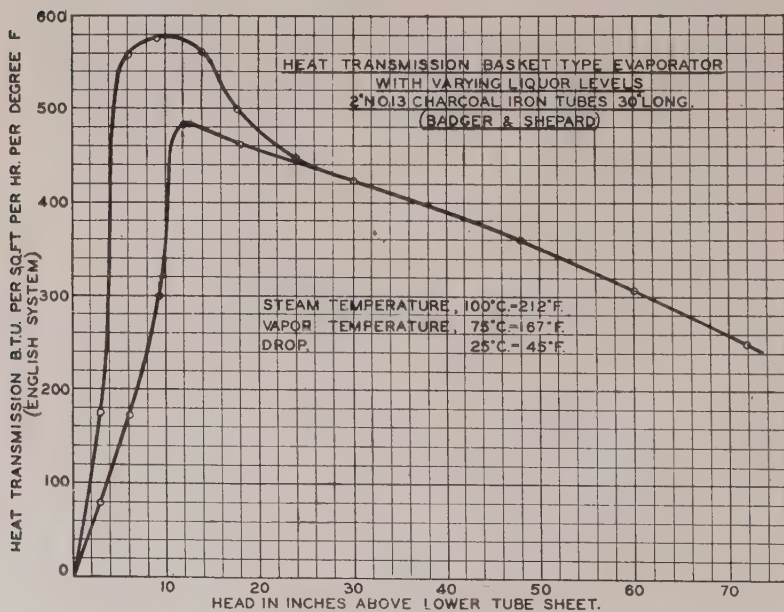


FIG. 9.

Badger and Shepard, *Am. Inst. Chem. Eng.*, Vol. XIII, pt. 1, p. 115, show the results of some experiments on the effect of temperature level, and of temperature difference on an experimental evaporator. These are given in Fig. 8 which has been platted on the English system, the original tests having been made in the metric system. An attempt was made to correct these results for hydrostatic head and steam temperature without much success.

The same writers, on page 144 of the same volume, show the results of their experiments on the effect of the level of the boiling liquid over the tubes of the evaporator, the so-called hydrostatic head effect. The diagram, Fig. 9, has been changed to the English system as before. The evaporator was of the basket type, and the maximum capacity was found to be when

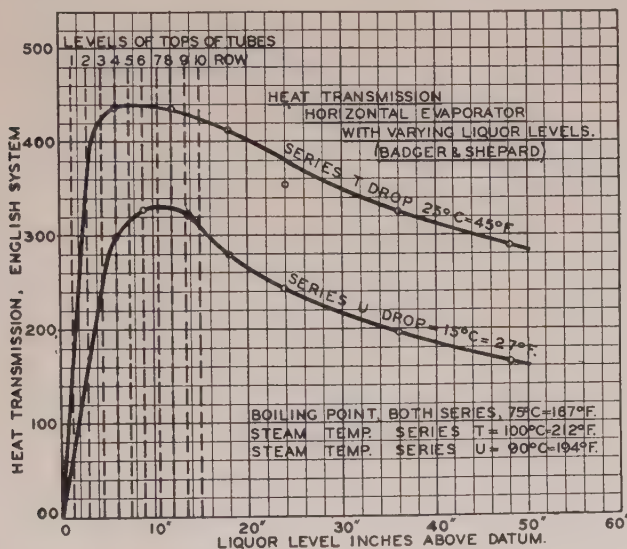


FIG. 10.

the liquor was a foot above the bottom tube sheet. This is of interest when compared with other data brought out in the various discussions in this book.

W. L. Badger has also published the results of tests of like character on a horizontal tubular evaporator, in the *Transactions of the American Institute of Chemical Engineers*, Volume XIII, part 2, page 148. These tests are given above in Fig. 10 in the English system.

We have collected a large amount of data on heat transmission, both experimental and on commercial machines. Extensive experiments are now under way at the Massachusetts Institute of Technology and it is hoped that in the not too distant future, the information will be generalized and become available.

Under present conditions, it is well to reiterate that the most practical way to determine what may be expected of an evaporator under specific conditions where data on similar work are not available, is to secure them on a similar machine, running under the same conditions as to liquor and vapor velocity, and with similar pressure and temperature differences, correcting for liquor viscosity as suggested in this chapter.

Chapter 6.

Basic Principles.

The basic principles of single and multiple effect evaporation are fairly simple and quite easy to understand. A brief discussion will bring out the essential points.

Any evaporating apparatus consists of two closed spaces separated from one another by a division made of thin metal in the form of tubes or plates of various shapes and proportions. In one space, steam is being condensed, giving up latent heat, and in the other, water or liquid at a lower temperature is being evaporated. On one side, therefore, steam is being condensed into water, and on the other side, water is being evaporated into steam.

The amount of heat given up by each pound of steam on condensation is that which corresponds to its latent heat at its particular pressure or temperature, and the amount of heat taken up by each pound of water evaporated is that which corresponds to its latent heat at its particular pressure or temperature. Two reservations are necessary.

When the steam condenses, giving up its latent heat, there is formed water at the same temperature as the original steam, and containing the sensible heat. Since the liquid boiling on the other side is cooler than this condensed steam, this condensate tends to cool to a temperature lower than that of the steam producing it. The amount of heat given up by each pound of steam, therefore, would be in excess of its latent heat by the amount of this condensate temperature reduction. In practice, the correction is so slight, that it is left out of account, for whereas condensate tends to be cooled by the colder liquid on the other side of the heating surface, on the other hand, it tends to remain at steam temperature, being in direct contact with steam, and having a relatively large surface exposed. It can therefore be said that for all practical purposes the amount of heat given up by steam on condensation in an evaporator is its latent heat.

The water or liquid as it is admitted into the liquor space of the evaporator may have a temperature corresponding to the pressure at which evaporation is taking place. Under these conditions, the amount of heat required to evaporate a pound of water will correspond approximately to the latent heat of the water at this temperature or pressure.

If, on the other hand, the water or liquid admitted into the evaporator has a temperature lower than that corresponding to the pressure at which evaporation is taking place, the heat required per pound of evaporation will be in excess of the latent heat of steam at this pressure or tempera-

ture, by the amount required to bring the water or liquid to the boiling point.

The converse of this proposition is also true. If the temperature of the water or liquid admitted into the evaporator is higher than that corresponding to the pressure at which evaporation is taking place, the heat required per pound of evaporation will be less than the latent heat of steam at this pressure or temperature by the amount of the liquor temperature in excess of the boiling point. This extra sensible heat is called "flash," or self evaporation.

It generally happens in evaporator work that the amount of liquid fed is considerably more than the evaporation. Here, if the temperature is above the boiling point, some of the water or liquor fed in will flash, making the amount of heat additionally needed per pound of evaporation still less. In the same way, if the temperature of the liquid fed is lower than the boiling point, the heat required per pound of evaporation will increase by the amount required to bring up the temperature of the total amount of liquid fed in up to its boiling point.

Thus it is necessary to analyze very carefully the exact conditions of operation before being able to determine the evaporation per pound of steam supplied. These conditions are subject to considerable fluctuation. Since the specific heat of the solution being evaporated may be considerably less than unity, it may even be necessary, when the variation is sufficient, to make a correction for this. Usually, such a correction can be omitted except where it would involve a considerable error, as it further complicates the calculations without materially affecting the results.

A few examples are to the point:

1) If 1 lb. of water is fed into an evaporator at 212° F., and evaporated at that temperature, the heat required will correspond to the latent heat of steam at that evaporation, or 970.4 B.t.u.'s. (Marks and Davis Steam Tables.)

2) If this pound of water enters at a temperature of 70° F., the amount of heat required for its evaporation will be 970.4 plus $(212 - 70 = 142)$, or 1111.4 B.t.u.'s.

3) If the temperature had been 250° F., the heat required for its complete evaporation at 212° F. would have been, 970.4 minus $(250 - 212 = 38)$, or 932.4 B.t.u.'s.

If two pounds of water had been fed to the evaporator for every pound of water evaporated in the above three cases, the heat requirements would have been as follows:

1. Heat required = L at $212 =$	970.4 B.t.u.'s
2. Heat required = L at $212 =$	970.4 B.t.u.'s
Plus liquor heating, $2 \times (212 - 70 = 142) =$	284.0 B.t.u.'s
Total	1,254.4 B.t.u.'s
3. Heat required = L at $212 =$	970.4 B.t.u.'s
Minus flash, $2 \times (250 - 212 = 38) =$	76.0 B.t.u.'s
Total	894.4 B.t.u.'s

Assuming that the steam on the other side of the heating surface was at 5 lbs. gauge, or 227° F., upon condensation, the heat given up by one pound of steam will be its latent heat at that pressure and temperature, or 960.7 B.t.u.'s, and the amount of steam per pound of evaporation in the above cases would have been as follows:

1. Steam per pound of evaporation, $\frac{970.4}{960.7} = \dots\dots\dots 1.010$
2. Steam per pound of evaporation, $\frac{1254.4}{960.7} = \dots\dots\dots 1.305$
3. Steam per pound of evaporation, $\frac{894.4}{960.7} = \dots\dots\dots .930$

The above is the basis of all evaporator calculations. When Norbert Rillieux first conceived the idea of multiple effect evaporation, he had no steam tables as we know them to-day, and thermodynamics was almost unknown as an applied science, so he formulated what is generally known as *Rillieux's First Principle*, which may be stated as follows:

"In a multiple effect, one pound of steam applied to the apparatus will evaporate as many pounds of water as there are bodies in the set." In other words, a single effect evaporates one pound of water per pound of steam; a double, two pounds of water per pound of steam; a triple, three, and so on.

In the light of the discussion above, it is evident that this method of calculation gives merely a very crude approximation, if we may call it such, whose correctness will vary greatly as the imposed conditions of operation are changed, the error being sometimes in one direction, and sometimes in the other.

It is rather surprising that to this day, many engineers, and many writers, still use Rillieux's First Principle in making their evaporator calculations. The only justification for so doing is that by this means, the calculations are greatly simplified, but it is undeniable that they are unsound, misleading, and incorrect. In Rillieux's days, it was the only rule available and as such had to be used.

In view of the fact that as the temperature goes down, the value of latent heat increases, it will be true, inasmuch as the temperature of the liquid being evaporated will always be less than that of the steam producing this evaporation, that even when liquor or water is being fed into the evaporator at the boiling point or lower, the evaporation per pound of steam will be less than unity.

For the heat to go through the dividing wall or heating surface, referred to in the beginning of this discussion, there must be a substantial difference between the temperature on the two sides. This is thoroughly discussed elsewhere.

In the examples given above, the figures were taken arbitrarily. There is nothing to prevent operating, for instance, with 10 lbs. on the steam side and 5 lbs. on the liquor side, or at any other pressures or temperatures, so long as the steam is hotter than the water or liquor. For that matter, neither is one limited to any particular difference of temperature or pres-

sure between the two sides of the heating surface, except by practical considerations. Further, there is nothing to stop the selection of a range of pressures below the atmospheric, such, for instance as from 0 gauge to a vacuum of 8 inches, or from 8 inches vacuum to 15 inches, or any other temperatures, vacua, or range of vacua or temperatures, except as limited by operating difficulties.

Therefore, the vapor leaving the liquid side of the evaporator in the original example can be taken and used as steam for another evaporator working on a range, for example, from 0 lbs. pressure to 8 inch vacuum, and in turn, the vapor of this body may also be used as steam for an evaporator operating with a vacuum of 15 inches on its liquor side, and so on. This, in brief, is multiple effect evaporation.

The conditions in each body must be analysed as suggested above to obtain the true results. It is evident at a glance, that as the number of evaporators or bodies in series are increased, the amount of steam per pound of total evaporation will decrease. One would therefore be tempted to use a large number of such bodies in order to reduce the steam consumption. There are certain limiting factors that determine the number of bodies permissible, the conditions of operation being given. This will be discussed elsewhere. Suffice it to say that the development has been gradual, and the progression from singles to doubles, to triples, etc., came about step by step, and that now, conditions permitting, there are series involving the use of as many as eight bodies, and under specially favorable conditions, twelve bodies have been used in a multiple effect.

Usually, it is necessary to design the apparatus to operate within a limited range of temperatures and pressures. The low end of the range is fixed by the highest vacuum obtainable commercially, and this depends generally upon the temperature of the injection water to the condenser. The high end of the range is fixed as a rule by the permissible pressure of exhaust steam, depending upon the design of the prime movers. Most multiple effects operate on exhaust steam, and if live steam is used instead, the upper temperature limit is determined by some other factor.

It must be noted here, that, in general, the evaporation per unit of heating surface depends upon the difference of temperature between its two sides. It is evident that a single effect under the same terminal conditions roughly evaporates about twice as much as each body of a double effect per unit of heating surface, or three times as much as a triple and so on. For a given amount of evaporation, the cost of the apparatus increases greatly with the number of bodies, almost in proportion to their number.

It is true, that the larger the number of bodies, the smaller the condensing equipment required for a given total load, and the lower the steam consumption, as stated above.

It is customary to feed the liquor to be evaporated in series through the various bodies, and the concentration is continuous and progressive, being maximum and ultimate in the last effect. Usually, the temperature of the liquor entering is lower than the boiling point in the first effect. Therefore, less than one pound of evaporation per pound of steam is ob-

tained at this point, and more than one pound of evaporation per pound of condensed vapor in all the other bodies, when the liquor enters the other bodies above the boiling point. The total ratio of pounds of water evaporated per pound of steam used will as a rule be less than the number of bodies. This depends to a great extent upon how much heating must be done in the first body before evaporation begins. When the initial temperature of the liquid being fed to the apparatus is higher than that prevailing in the liquor space of the first effect, this ratio of water evaporated per pound of steam used will be greater than the number of bodies, but this condition seldom exists.

There is given in the next chapter a full discussion of a logical method of analysing the work of evaporators.

The series feed referred to above is not universal. Sometimes, for special reasons, the liquor is pumped through the bodies in the reverse order, entering the last body, and leaving the first. This is called counter current operation. It has a number of applications, and will be discussed in detail at another point.

Sometimes in certain industries, it is customary to feed each body of the multiple effect, concentration taking place in each body independently of the others. This is called parallel feed operation, and will be discussed also.

Chapter 7.

The Heat Balance.

It was mentioned under the discussion of Basic Principles that the old method of making evaporator calculations devised by Rillieux was not satisfactory, particularly in the statement of his First Principle which was "that in a multiple effect, one pound of steam applied to the apparatus will evaporate as many pounds of water as there are bodies in the set. This

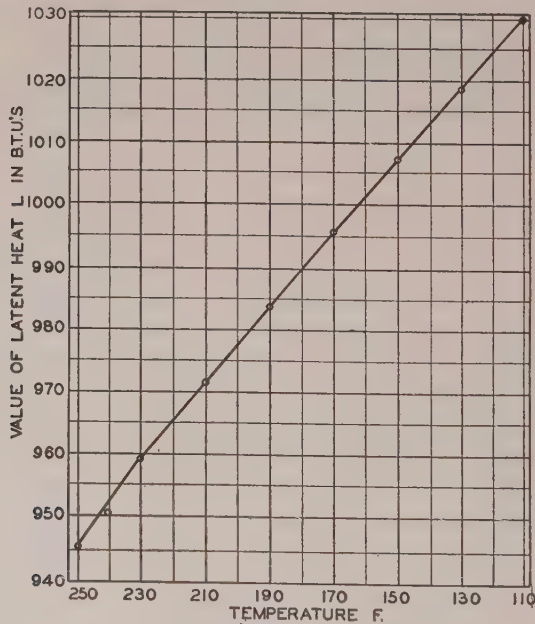


FIG. 1.—Variation of Latent Heat with Temperature.

is not exactly true under any condition, and far from being true under average operating conditions. It was originally based on the assumption that for all pressures existing in the various bodies the latent heat of steam, L , was constant and to all intents and purposes 1000 B.t.u. per pound. That this assumption is not justified is shown by the curve in Fig. 1, which shows the varying values of this latent heat within the range of temperatures commonly encountered in evaporator practice.

Another error which is quite prevalent is the fact that the liquid fed to the evaporator passes usually through the various bodies in series, from the high-temperature end to the low-temperature end, and as it goes from a hot to a cooler body a "flash," or self-evaporation, takes place, bringing the feed down to the temperature of the body into which it is entering. When this evaporation is taken into account, it has a very material effect on the balances of heat flow, depending upon the weight per unit of time and the range of temperature fall.

Temperature of Feed Liquor Most Important. By far the largest variable in estimating the amount of evaporation per pound of steam is the temperature of the liquid as it enters the multiple effect, as compared with the existing boiling temperature in the first body. Obviously if the liquid must be heated to the boiling temperature, steam must be used up which would otherwise be available for evaporation. On the other hand, if the liquid temperature is higher than that in the first body, a "flash" takes place and actually more than a pound of water will be evaporated per pound of steam. The same condition exists as a rule in the subsequent bodies of a multiple, since there is always a "flash," and the evaporation will be greater than the amount of vapor condensed on the steam side. Therefore, in making calculations of the evaporation possible in a given effect it is a matter of balancing the latent heat available from the condensation of steam in the calandria against the latent heat represented by evaporation on the liquor side, to which must be added the heat required to bring the liquor to a boil if it is too cold, or from which must be subtracted the heat represented by "flash" if it is too hot.

The Use and Calculation of Heat Balances. The only way to determine accurately the operation of an evaporator is by making a thorough study of the heat transferred, with the aid of a heat and liquor flow sheet, or heat balance, in which all the steps and their interrelations are shown. When conditions are known, it is not difficult, and this method of analysis permits a comprehensive grasp of relatively complicated evaporator problems. This will be illustrated by an analysis of one of the most interesting problems in evaporation—namely, the determination of the most advisable cycle of feed, governed by the consideration of steam economy alone.

Assume a triple effect, whose bodies are numbered 1, 2, 3, beginning at the steam end and finishing at the condenser end. The question to be discussed is this:

All conditions being precisely the same, with the exception of the temperature of the liquid entering, which is variable, shall the feed go through the three bodies in series, 1, 2, 3, commonly known as "parallel current," or 3, 2, 1, "counter current," or shall each body be fed independently?

There are many conditions aside from heat economy which preclude the possibility of using some of these cycles, but in this case it will be assumed that heat economy is the only determining factor.

In order to avoid unnecessary complications which would cloud the issue, assume arbitrarily the following conditions, which are not what

are found in practice, but which will serve well enough for this discussion, which should not be complicated by the introduction of some forty-odd variables actually entering in evaporator work:

Liquid to be treated per hour.....	100,000 lbs.
Evaporation required	70,000 lbs.
Evaporation in triple effect.	
Initial feed temperature.....	Unknown and variable
Specific heat of solution.....	Unity
Viscosity loss	Negligible
Boiling temperature rise.....	Negligible

	Steam Belt 1	Steam Belt 2	Steam Belt 3	Vapor Belt 3
Press and vacuum.....	5.0 lbs.	2 in vac.	14 in vac.	26 in vac.
Temperature	227°	209°	182°	125°
Latent heat	960.7 B.t.u.	972.2 B.t.u.	988.7 B.t.u.	1,021.6 B.t.u.

Complete heat balances have been calculated for "parallel current" operation, also for "counter current," and "parallel feed" for initial feed temperatures of 50°, 100°, 150°, 200° and 250°. It would take up too much space to give them all completely here. Therefore abstracts are given for each cycle, giving the information required to compare per-

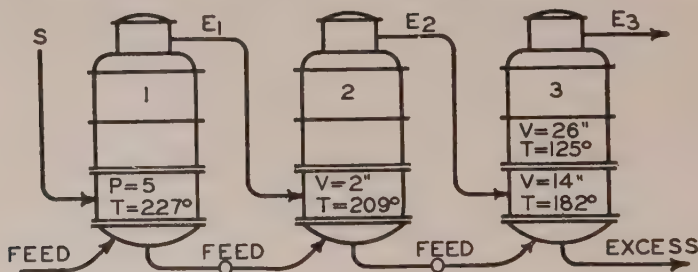


FIG. 2.—Diagrammatic Sketch of Parallel Current Operation.

formances. One of these heat balances for each cycle will be given, calculated for a feed temperature of 150°, as representing probable average conditions encountered in practice.

Calculation of "Parallel Current" Heat Balance. In the calculation given in Table I, begin by admitting into the steam belt of the first effect 27,800 lb.* of steam per hour at a pressure of 5 lbs. gauge, corresponding to a temperature of 227°, which upon condensation gives up its latent heat 960.7 B.t.u. per pound, making a total of 26,700,000 B.t.u. liberated.

This is transmitted to the liquor through the heating surface, part of it being used to bring the feed temperature to the boiling point, in this case, 209°. The initial temperature being 150°, the necessary rise is 209°—150°, or 59°. Inasmuch as the entire feed enters the first body, we must

* This figure is reached by a process of trial and error which necessitated calculating the complete heat balance after assuming this initial figure. Three or four calculations are usually sufficient to arrive at a figure which will check out within operating limits of accuracy.

TABLE I.
PARALLEL CURRENT HEAT BALANCE.

Specifications:		Feed Temperature = 150° F.	Heat	Liquor
1.	Steam, 27,800 lbs. at 5 lbs. = $27,800 \times 960.7 =$		26,700,000	100,000
	Deduct for heating 100,000 $\times (209 - 150 = 59) =$		5,900,000	
	Available for evaporation.....		20,800,000	
	L at 209 = 972.2. $E_1 = \frac{20,800,000}{972.2} =$			21,400
				78,600
2.	Vapor ex. 1.....		20,800,000	
	Add flash, 78,600 $\times (209 - 182 = 27) =$		2,120,000	
	Available for evaporation.....		22,920,000	
	L at 182 = 988.7. $E_2 = \frac{22,920,000}{988.7} =$			23,200
				55,400
3.	Vapor ex. 2.....		22,920,000	
	Add flash, 55,400 $\times (182 - 125 = 57) =$		3,160,000	
	Available for evaporation.....		26,080,000	
	L at 125 = 1,021.6. $E_3 = \frac{26,080,000}{1,021.6} =$			25,500
				Excess..... 29,900

heat it all, 100,000 lbs., through this rise of 59°, thus absorbing 5,900,000 B.t.u. The heat left available for evaporation is therefore 26,700,000 — 5,900,000, or 20,800,000 B.t.u. The latent heat at the temperature of ebullition, 209° in the first effect, is 972.2 B.t.u. per pound. The evaporation in the first effect will therefore be 20,800,000 divided by 972.2, or 21,400 lbs., and the amount of feed transferred from the first to the second effect will be the amount of the original feed, 100,000 lbs. less this evaporation, 21,400, or 78,600.

For the second effect, heat is obtained from two sources*: First, the vapor from the first effect, which upon condensation gives up its latent heat, shown above to be 20,800,000 B.t.u.; and second, the heat obtained from the flash or self-evaporation taking place when the feed leaves the first body at a temperature of 209° and enters the second at 182°, thus giving up 27 B.t.u. per pound. The amount of this liquor being 78,600, as shown above, the total heat liberated in this way will be 78,600 times 27, or 2,120,000 B.t.u., making a total of 22,920,000 available for evaporation in the second body. The latent heat at this point at the existing temperature of 182° is 988.7 B.t.u. per pound, and the corresponding evaporation will be 22,920,000 divided by 988.7, or 23,200, and the liquor transferred from the second to the third effect will be the amount which entered this body, 78,600, less this evaporation, 23,200, or 55,400 lbs. per hour.

* Condensates are removed from each effect, and not carried to succeeding ones.

A similar exchange takes place in the third effect. First the vapor from the second body condenses, giving up its latent heat, shown above to be 22,920,000 B.t.u., and the feed from the second flashes down from its original temperature of 182° to that of the third effect, 125° , thus giving up 57 B.t.u. per pound. The quantity being 55,400 lbs., the heat thus liberated is 3,160,000 B.t.u., which when added to the above gives a total of 26,080,000 available for evaporation in the third effect. The latent heat of steam at the third body temperature, 125° , is 1021.6 B.t.u. per pound, and the evaporation here is therefore 25,500 lbs. per hour, which when deducted from the feed into the third effect, 55,400 lbs., leaves 29,900 lbs. which is removed from the last body as concentrated liquor. This

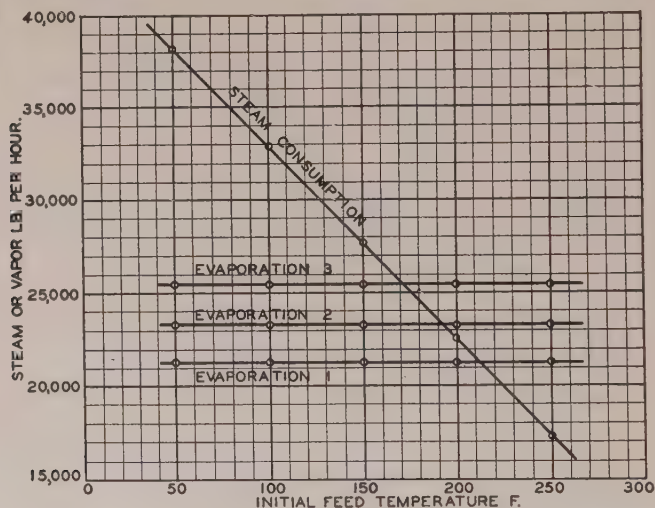


FIG. 3.—Triple Effect Performance with Parallel Current Operation at Varying Temperatures.

compares closely with the 30,000 which we should obtain. The variation is negligible.

Fig. 3 is a diagram plotted from a summary of the five heat balances calculated for various feed temperatures for "parallel current" operation, which represents the cycle used most extensively. The reader must bear in mind that in all these cases the total evaporation for the triple effect is the same—namely, 70,000 lbs. per hour—and the total feed is the same, 100,000 lbs. per hour. The only variation is in the temperature of this feed. The steam supply is then varied to give 70,000 lbs. evaporation per hour in all cases. With the "parallel current" cycle the evaporation in each body does not vary as the temperature of the feed changes. By referring again to the heat balance shown in Table I the reason for this can be readily seen. In order to accomplish the evaporation required, the first body must produce 21,400 lbs. of vapor. In order to do this, there must

be available for evaporation 20,800,000 B.t.u. If the liquor enters at the boiling point, then we must put into the steam belt of the first effect 20,800,000 B.t.u. by condensing the corresponding amount of steam. If the feed is below the boiling point, we must add sufficient heat to bring it to a boil, increasing the steam consumption to that extent. If, on the other hand, liquor enters above the boiling point, a corresponding flash takes place, liberating a certain amount of heat, thus decreasing the amount of steam required to make up the necessary 20,800,000 B.t.u. as vapor leaving the first body. The evaporation is greater in the second than in the first effect, and greater in the third than in the second, for reasons discussed previously. The temperature of the concentrated liquid is 125° F. It must be noted that the steam consumption varies within rather wide limits as the feed temperature changes, a fact well worth remembering.

"Counter Current" Heat Balance. We shall now go over the calculations for the "counter current" heat balance, summarized in Table II. In

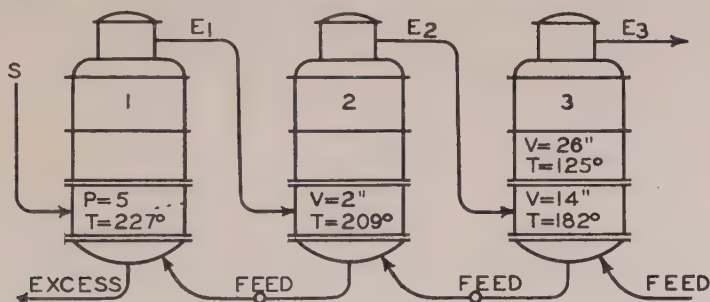


FIG. 4.—Diagrammatic Sketch of Counter Current Operation.

this case it is easier to work backward, inasmuch as the feed flows backward.

Begin by feeding into the third effect 100,000 lbs. of liquor at the given temperature of 150° . By assuming 22,880 lbs. is evaporated at this point. The latent heat of steam at the existing temperature, 125° , is 1021.6 B.t.u. per pound, and the heat in this vapor is therefore 22,880 times 1,021.6, or 23,360,000. Part of this heat is derived from the flash or self-evaporation of the feed, 100,000 lbs., which enters at 150° , dropping down to the boiling point, 125° , and giving up 25 B.t.u. per pound, or a total of 2,500,000 B.t.u. Inasmuch as the vapor contains 23,360,000 B.t.u., as stated above, the difference, 20,860,000 B.t.u., must be made up by the condensation in the steam belt of the third effect of vapors from the second at a temperature of 182° , having a latent heat of 988.7 B.t.u. per pound. The amount of vapor thus condensed will therefore be 20,860,000 divided by 988.7, or 21,100 lbs., which of course represents the evaporation in the second effect.

The feed transferred from the third to the second effect is the total feed, 100,000 lbs., less the evaporation in the third effect, shown above by assumption to be 22,880 lbs., giving 77,120.

TABLE II.
COUNTER CURRENT HEAT BALANCE.

Specifications:		Feed Temperature = 150° F.		Heat	Liquor
3.	Vapor ex. 3 = 22,880 lbs. at 26 in. = $22,880 \times 1,021.6 =$			23,360,000	100,000
	Deduct flash, $100,000 \times (150 - 125 = 25) = \dots\dots\dots$			2,500,000	$E_2 = 22,880$
	Heat required S.B. 3.....			20,860,000	77,120
	L at 182° = 988.7. $E_2 = \frac{20,860,000}{988.7} = \dots\dots\dots$				$E_2 = 21,100$
2.	Vapor ex. 2.....			20,860,000	56,020
	Add for heating, $77,120 \times (182 - 125 = 57) = \dots\dots\dots$			4,400,000	
	Heat required S.B. 2.....			25,260,000	
	L at 209 = 972.2. $E_1 = \frac{25,260,000}{972.2} = \dots\dots\dots$				$E_1 = 26,000$
			Excess.....		30,020
1.	Vapor ex. 1.....			25,260,000	
	Add for heating $56,020 \times (209 - 182 = 27) = \dots\dots\dots$			1,511,000	
	Heat required S.B. 1.....			26,771,000	
	L at 227 deg. = 960.7. Steam = $\frac{26,771,000}{960.7} = \dots\dots\dots$				27,850

In the second effect, there must be supplied not only the latent heat contained in the vapors going to the third, shown above to be 20,860,000 B.t.u., but in addition there must be heated up the feed, 77,120 lbs., transferred from the third effect at a temperature of 125° to the second at a temperature of 182°, a rise of 57°, giving 77,120 times 57, or 4,400,000 B.t.u., which when added to the latent heat of the vapor above, 20,860,000, gives 25,260,000 B.t.u. as the total heat requirements in the steam belt of the second body. Now the temperature of this steam is 209° and the latent heat for this is 972.2°. The amount of steam condensed in the steam belt of the second effect will therefore be 25,260,000 divided by 972.2, or 26,000 lbs., which is also the evaporation in the first effect.

The feed transferred from the second to the first effect will be the amount fed into the second from the third, 77,120, less the evaporation in the second, shown above to be 21,100, or 56,020 lbs.

In the first effect, heat enough must be provided to correspond to that contained in the vapors therefrom, given above as 25,260,000 B.t.u., plus the heat required to bring up the feed, 56,020 lbs., from the temperature of the second body, 182°, to the temperature of the first, 209°, a rise of 27°, thus making up 56,020 times 27, or 1,511,000 B.t.u., which, when added to the heat in the vapors from this effect, 25,260,000, gives a total of 26,771,000 B.t.u. This heat must be supplied by condensing steam in the first steam belt at a temperature of 227°, at which the latent heat of steam is 960.7 B.t.u. per pound. The quantity of steam required is therefore 26,771,000 divided by 960.7, or 27,850 lb. per hour.

The concentrated liquor discharged from the first effect will be the

amounted from the second to the first, 56,020, minus the evaporation in the first, shown above to be 26,000, giving us 30,020. In this case also, by premise, this quantity should be 30,000.

In Fig. 5 the heat balances are plotted graphically for different feed temperatures. The feed enters the third body, and is pumped from there to the second and then to the first, going in opposite direction to the steam, hence the name. The evaporation in each body varies considerably as the initial feed temperature changes. The steam consumption as well as the evaporation in the first and second effect decreases as the feed temperature increases. The evaporation in the third effect, however, does just the opposite. This is readily understood by studying the heat balance.

When the feed temperature is below 125° , as it enters the last body, it must be heated up before it evaporates, thereby increasing the heat

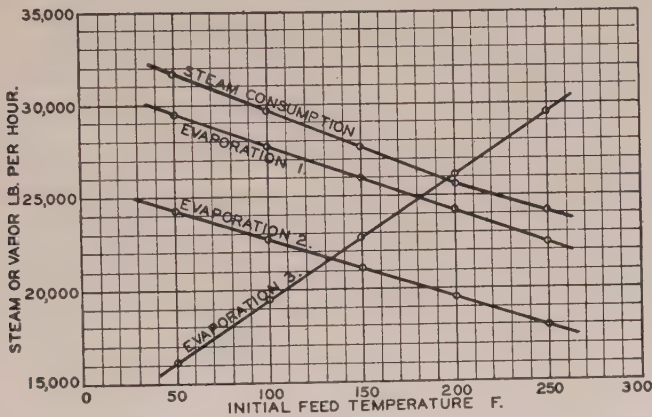


FIG. 5.—Triple Effect Performance with Counter Current Operation at Varying Temperatures.

requirements in the steam belt. When this temperature, however, is higher than 125° , a flash takes place, and this vapor is not used in multiple, but goes directly to the condenser. Thus, in the case of feed at 250° , there is 12,500,000 B.t.u. liberated as flash and 17,750,000 of transmitted heat, making a total of 30,250,000 B.t.u. in vapor to the condenser. With temperature of feed at 50° F., just the reverse takes place. There is 24,050,000 of transmitted heat, of which 7,500,000 is used in bringing the liquid to a boil and 16,550,000 represented as vapor.

It is to be noted that with counter current operation there is a flash only in the last body and only when the temperature of the entering feed is above 125° F., the boiling point. Except where there is a flash, each preceding body evaporates more than the succeeding one, for the liquid must be heated in each case before it evaporates, thus increasing the heat requirements of the second body over the third and the first over the second.

Generally speaking, the load on the condenser (the evaporation in the last effect) is much less than with "parallel current" operation, especially at low feed temperatures. It is also to be noted that the concentrated liquid leaves the first effect at a temperature of 209° , as against 125° for parallel current operation, a fact well worth considering if subsequent heating operations are required after the concentrated material leaves the apparatus. When viscous materials are to be evaporated, it is very advantageous to operate "counter current," as, in view of the high temperatures at high concentration, the annoying effect of high viscosity are very much reduced.

Parallel Feed Heat Balance. Parallel feed heat balance is calculated with initial feed at 150° F., as before. Begin by admitting 26,700 lbs. of steam into the calandria of the first effect, obtaining for useful work its latent heat at the existing temperature 227° , which is 960.7 B.t.u. per pound. The heat thus liberated is therefore 25,680,000 B.t.u. Each

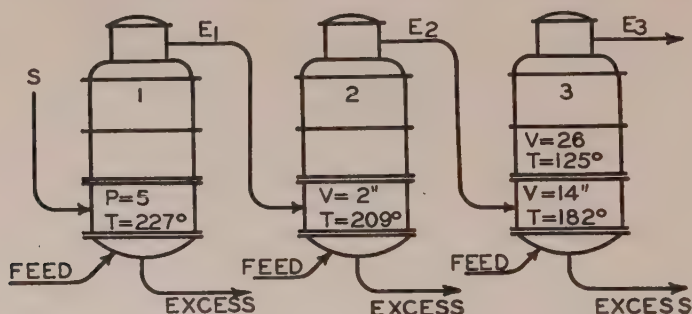


FIG. 6.—Diagrammatic Sketch of Parallel Feed Operation.

pound evaporated will require in the first place an amount of heat equal to the latent heat at the temperature existing in the first effect, 209° , or 972.2 B.t.u. This pound to be evaporated represents only 70 per cent. of the corresponding liquor fed, for, by premise, there must be evaporated 70 per cent. by weight in each body. This liquor must be heated from its initial temperature of 150° to the temperature of the first body, 209° , a range of 59° ; therefore the amount of heat required for this operation will be 59 divided by 0.7, or 84.3 B.t.u. per pound of evaporation, which when added to the latent heat above, 972.2, gives a total of 1056.5. The evaporation in the first effect will therefore be the total heat available given above as 25,680,000 divided by this 1056.5, or 24,300 lbs. per hour.

In the second effect, when this vapor condenses, it gives up its latent heat only, or 972.2 B.t.u. per pound. The available heat at this point will therefore be 24,300 times 972.2, or 23,600,000 B.t.u.

The heat required per pound of evaporation will be the latent heat at the temperature of the second effect, 182° , which is 988.7, to which must be added the necessary amount to bring up the feed from 150° to 182° , a rise of 32° . Here again the evaporation represents only 70 per cent. of the

TABLE III.
PARALLEL FEED HEAT BALANCE.

Feed Temperature = 150° F.

	Heat	Evapo- ration
Specifications:		
1. Heat available steam belt No. 1.		
26,700 lbs. at 5 lbs. = $26,700 \times 960.7 =$	25,680,000	
Heat required per lb. evap. as follows:		
L at 209° = 972.2		
Add for heating (70 per cent. evap.)		
$\frac{100}{70} \times (209 - 150 = 59) =$	84.3	
Total	1,056.5	
Evaporation 1 = $\frac{25,680,000}{1,056.5} =$		24,300
2. Heat available steam belt No. 2.		
24,300 lbs. at 209° = $24,300 \times 972.2 =$	23,600,000	
Heat required per lb. evap. as follows:		
L at 182° = 988.7		
Add for heating (70 per cent. evap.)		
$\frac{100}{70} \times (182 - 150 = 32) =$	45.7	
Total	1,034.4	
Evaporation 2 = $\frac{23,600,000}{1,034.4} =$		22,800
3. Heat available steam belt No. 3.		
22,800 lbs. at 182° = $22,800 \times 988.7 =$	22,550,000	
Heat required per lb. evap. as follows:		
L at 125° = 1,021.6		
Deduct for flash (70 per cent. evap.)		
$\frac{100}{70} \times (150 - 125 = 25) =$	35.7	
Total	985.9	
Evaporation 3 = $\frac{22,550,000}{985.9} =$		22,800
		69,900

liquor fed in, and this heating will therefore require 32 divided by 0.7, or 45.7, which when added to the latent heat above, 988.7, gives us 1034.4 B.t.u. required for each pound of evaporation as before. The total heat available was shown above to be 23,600,000 B.t.u., and the evaporation is therefore 22,800 lbs. per hour.

As in the case of the second effect, when this vapor condenses in the third steam belt it gives up only its latent heat, stated above as 988.7 B.t.u. per pound. The heat available for evaporation in the third effect is therefore 22,800 times 988.7, or 22,550,000 B.t.u.

The heat required per pound of evaporation will be, as before, the latent heat of steam at the temperature of the third effect, 125°, which is 1021.6. The correction for the heating effect in this case is negative, for the feed enters at a temperature of 150°, which is 25° higher than that of the third effect. A flash will therefore result liberating 25 divided

by 0.7, or 35.7 B.t.u. per pound of evaporation. Deducting this from the latent heat of steam for this temperature, given above as 1021.6, we have 985.9 B.t.u. as the amount of heat required for each pound of evaporation in the last effect. The available heat in vapors from the second body is given above as 22,550,000 B.t.u. The evaporation is therefore 22,800.

The total evaporation for the entire equipment is as follows:

First effect	24,300
Second effect	22,800
Third effect	22,800
Total	69,900

By premise this amount should be 70,000, and the discrepancy is therefore 100 lbs. per hour.

In the case of "parallel feed" operation each body of the triple is fed independently, and the concentrated liquor removed separately from each.

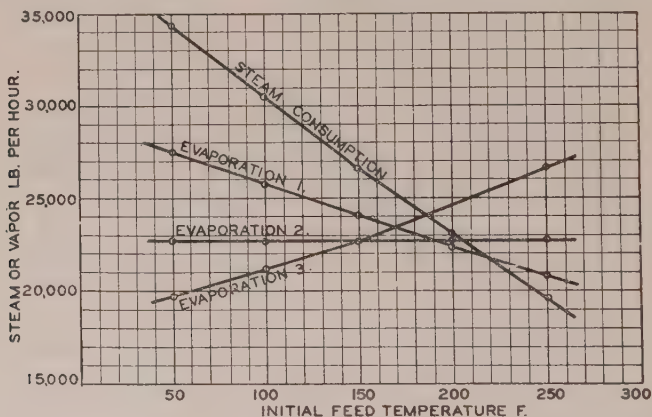


FIG. 7.—Triple Effect Performance with Parallel Feed at Varying Temperatures.

In performance, this cycle lies roughly between the other two. It has only very few limited applications in practice, and is discussed here more as a matter of interest. Perhaps its most important use is in the salt industry, where one starts with saturated brine and the same concentration exists in each body. It is true that for a certain limited range it is more economical than either of the other two cycles, in this case, from feed at 125° to feed at 175°. According to the temperature of the feed, there may be flashing in all bodies, heating in all bodies, or heating in some and flash in others. The steam consumption decreases as the temperature of the feed increases; so does the evaporation in the first effect. The evaporation in the second effect is practically constant, and in the third increases as the feed temperature increases.

THE HEAT BALANCE

Conclusions.

Variation in Steam Consumption in Various Cycles. Fig. 8 shows a comparison of the steam consumption of the triple effect with varying

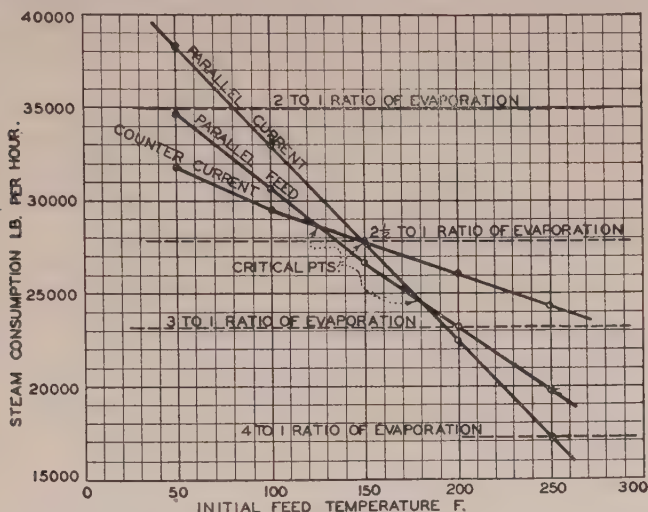


FIG. 8.—Comparative Steam Consumption for Various Cycles.

feed temperatures, operated according to the various cycles suggested above. The crossing point of the parallel and counter current curves is at a temperature of 150° F. Our readers are particularly warned

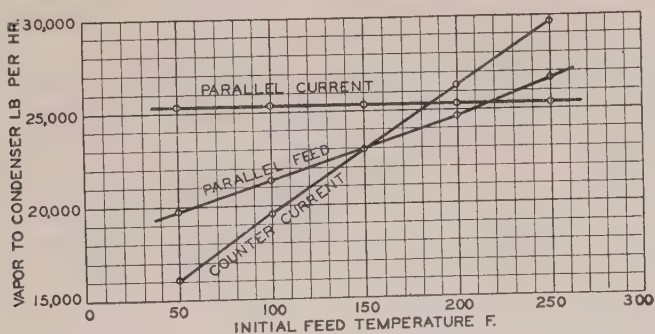


FIG. 9.—Condenser Load under Varying Conditions.

that this critical point applies only for our assumed conditions, and would be different if these conditions are changed. Therefore, under the prescribed conditions, the "parallel current" method of operation is more

economical if the temperature of feed is above 150° F., and the "counter current" if this temperature is less than 150° .

Condenser Load Variations. Fig. 9 indicates the varying load on the condenser, which of course is the same as the evaporation on the last effect. This is worked out for the three cycles for all temperatures. There is a decided advantage here in favor of the "counter current" cycle at lower feed temperatures.

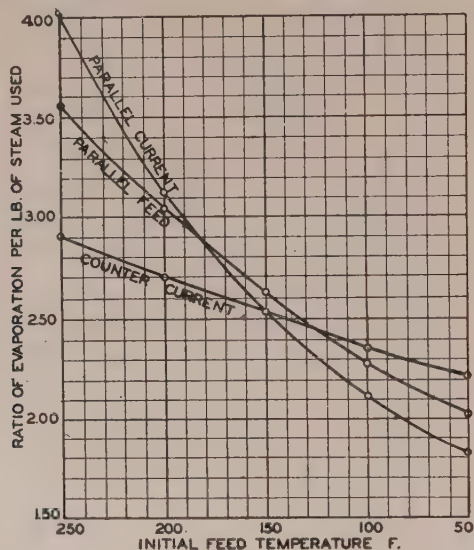


FIG. 10.—Evaporation per Pound of Steam under Varying Conditions.

The amount of evaporation in every case is the same by premise for the whole equipment—namely, 70,000 lbs. per hour. If this figure be divided by the steam consumption in any particular case, there is given the ratio of evaporation in pounds of evaporation per pound of steam used in the equipment. These data are worked out as shown in the accompanying Fig. 10.

In view of the discussion at the beginning, it is of interest to note the wide divergence from accepted information. These curves also bring out very forcibly the importance of keeping the feed temperature as hot as possible, and of keeping it so when once attained, if steam economy is of any consequence.

Chapter 8.

Practical Use of Heat Balance.

In the previous chapter, the proper method of making up heat balances for evaporators has been discussed for various cycles identified as parallel current, counter current, and parallel feed. A full discussion will now be given of the facts to be considered in determining the conditions of operation for a given problem, finally establishing the distribution of temperatures in the selected apparatus, and working out a heat balance to fulfil these requirements.

The most disconcerting factor to the beginner in this work is the large number of variables which have to be considered. When all the characteristics of the solution to be evaporated are well known, the process is laborious, but not difficult, and experience and practice will give a good indication of what may be expected. It is well to state here that the most careful theoretical calculations are very often completely upset by minor practical considerations. Such theoretical work, however, is invaluable in that it sets the limits within which practice must be maintained.

The generally accepted method of arriving at the physical characteristics of a given solution is by laboratory experiment. The experience of the author extending over many years of work and study on evaporation has shown that such small scale experiments may prove very misleading. This does not mean that it is not possible to determine certain general characteristics, such as the rise in boiling points for various densities, viscosity, surface tension, etc.

The main stumbling block is the determination of the coefficient of heat transmission which may be expected in the actual full-sized apparatus, working under practical conditions. Many factors affect this coefficient, the most important of which is the most elusive, namely, the effect of fouling and scaling of the heating surfaces. This is discussed in Chapter 5.

As to the effect of viscosity on heat transmission in an evaporator, there is not a great deal of direct information. This has been mentioned previously in the chapter on Factors Affecting the Coefficient of Heat Transmission. Experiments conducted at the Massachusetts Institute of Technology, noted elsewhere in this book, show that the coefficient of heat transmission is inversely proportional to a power function of the viscosity in a purely heating operation. It can therefore be assumed that evaporation would be affected in very much the same way, and there is no question but that the effect is considerable and should be taken into account.

The rise in the temperature of the boiling solution over and above the water vapor saturation temperature corresponding to the pressure at which

ebullition takes place can be very easily determined for practical purposes in the laboratory. The test can be made simply by boiling solutions at various densities which cover the range of contemplated evaporator operation, and noting the difference between the observed temperature of the liquid and the temperature of distilled water vapors at the same pressures. Then, knowing the probable density in each body of the multiple effect, additional allowance must be made in the temperature drop of each to correspond to this loss. For many standard solutions, this information has been determined and tabulated, and can be found in handbooks. (Refer to the appendix for such tabulations.)

The surface tension determination, which is usually an index of the foaming properties of the solution will be used to determine the type of evaporator to use. If the material has a marked tendency to foam, it is advisable to use long tube evaporators with very rapid vapor circulation. It has been well established that foam does not occur in apparatus of this kind if the velocity of travel in the tubes is more than a certain minimum, depending on the surface tension, in the neighborhood of 12 feet per second.

If the solution is of delicate nature, and easily affected by heat, it is necessary to select an evaporator having a very small volume of liquor in circuit, usually of the long tube type, and also, in addition, it is advisable to maintain low boiling temperatures by operation perhaps in single effect under high vacuum, and, in extreme cases, by using steam under partial vacuum in the calandria. Excellent results have been obtained by this procedure in concentrating such materials as photographic gelatine.

Different types of evaporators have different coefficients of heat transmission, and this fact also must be taken into consideration in the calculations. This will be readily appreciated by referring to the curves showing the effect of steam temperature on heat transmission shown in Chapter 5 referred to previously.

Another point which it is well to determine as data for estimating a correct heat balance is the latent heat of solution of the material to be concentrated at various densities. Usually, for ordinary solutions, this factor can be neglected. For such materials as caustic soda, glycerine, and others, lack of consideration would result in a material discrepancy. This latent heat of solution, when sufficient to be considered is added to the latent heat of steam to be evaporated in each body for the particular concentration existing therein, and the work then calculated on this new basis.

Number of Bodies to be Used. In the determination of the number of bodies to be used in the multiple effect under consideration, not only must the above factors be considered, but others as well, as the following discussion will show.

The total range of temperature available must be considered. This is determined usually by the steam pressure on the one hand, and the commercially obtainable vacuum on the other. This vacuum must not be taken at its best, but under the worst seasonable conditions, which generally resolves itself into a question of the temperature of the condensing

water available. Furthermore, it must be remembered that as the vacuum is raised beyond a certain point, it becomes more and more expensive to maintain, and it is very seldom practical and permissible to figure on more than 27 in. year in and year out. In tropical or semi-tropical climates, it is better to keep under 26 in., as the water is likely to reach a temperature of 90°. (Refer to Chapter 18 on Condensers.)

Usually, the upper limits of temperature are determined by the pressure of exhaust steam that can be conveniently carried by the engines or turbines in the plant. This will vary from 5 lbs. gauge, to 20 lbs. gauge, according to the plant. The corresponding pressure temperatures must be corrected for the barometer if the elevation is over 1000 feet.

Sometimes the quantity of exhaust steam is negligible, in which case, the live steam pressure is the determining factor. Under such operating conditions, this exhaust can be turned into the second or third body where the pressures correspond.

The cost of steam as compared with the cost of the equipment must be taken into account. If there is a large supply of exhaust steam which otherwise would be wasted, it would be an error to reduce the steam consumption below the available steam supply by increasing the number of effects, a profitless expenditure. One must also bear in mind that the more the number of bodies, not only does the equipment increase in cost, but the complications as well, and it is often good judgment to sacrifice some steam economy for the sake of simplicity and ease of operation, and low first cost.

Having fixed the total temperature fall between the steam and the condenser as indicated above, it is now in order to settle the question of the number of bodies in the set.

Knowing the initial and final density of the solution under consideration, and having determined the rise in boiling points, for the range of operation, a rough calculation will give the probable densities for a double, triple, quadruple, quintuple, and sextuple effect. The sum of the probable temperature losses can be deducted from the total available temperature range, leaving the difference as a working drop. Experience and study will indicate the minimum amount of this net temperature drop advisable for each body for practical operation.

Generally speaking, the higher the viscosity the more temperature difference will be required on account of the reduced coefficient of heat transmission. Also, the lower the steam pressure, the lower the coefficient of heat transmission, and the greater the temperature difference required. Fig. 2 gives a curve representing in a general way the trend of the coefficient of heat transmission as the temperature of steam changes. (Taken from *Louisiana Bulletin* 149, E. W. Kerr.) The data on this plot were determined experimentally on laboratory evaporators of small size, and different designs, operated on distilled water and with clean surfaces. There are considerable variations in the different evaporators tested, but in general, the trend is all the same, and seems to show that at 227° F., the heat transmission is about twice what it is at 160° F., so that from this standpoint alone, it can be said that the net drop in the

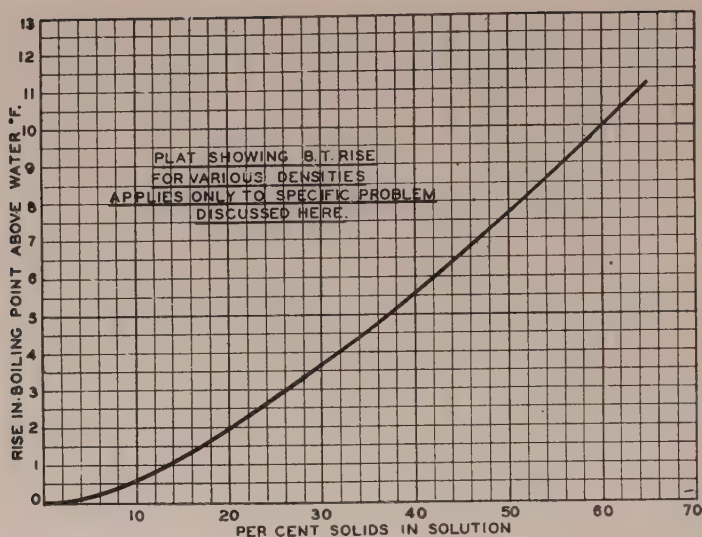


FIG. 1.

last effect must be about twice what it is in the first. This must be further increased to overcome the effect of viscosity, inasmuch as parallel current operation is contemplated.

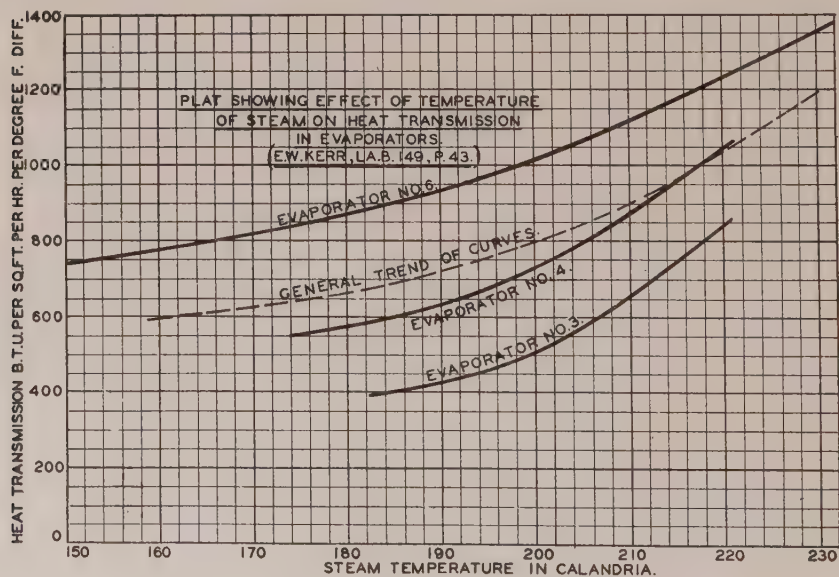


FIG. 2.

The losses of temperature due to hydrostatic head will be greater with low pressures than with high pressures, since a given increment of pressure at the low end of the scale corresponds to a much larger temperature increment, as will be readily seen by referring to the steam table in the Appendix. This subject of hydrostatic head losses is discussed in the chapter on Factors Affecting the Coefficient of Heat Transmission.

With low viscosities, it is possible to obtain high efficiency at atmospheric pressures with a total temperature fall on one body of as low as 8° to 10° F. Under high vacuum, it is necessary to increase the net drop to about 18° to 20°, and with moderate viscosities, such as for example the last bodies of multiple effects operating on sugar solutions, this must be increased to about 35° F. as a minimum net drop.

When the evaporators are perfectly clean, it is possible to do much better, but it is believed that the above represents average work that can be expected from apparatus in good condition and properly designed.

A concrete example will illustrate the above discussion.

A given plant has 100,000 lbs. of solution per hour to concentrate. The initial solid contents of this solution is 15 per cent. by weight, and it must be brought up to 60 per cent. solids by weight. The initial temperature is 180° F. Steam is available at 8 lbs. gauge, and the temperature of water for the condenser is such that it is practical to maintain a 26-in. vacuum, the barometer being 30 in. The boiling temperatures of this liquid is represented by Fig. 1.

LOG OF EVAPORATIONS AND DENSITIES IN VARIOUS INDIVIDUAL BODIES OF MULTIPLE EFFECTS.

	Single		Double		Triple		Quad.		Quint.		Sext.	
	L. & V.	Dens.	L. & V.	Dens.	L. & V.	Dens.	L. & V.	Dens.	L. & V.	Dens.	L. & V.	Dens.
Feed	100,000	15	100,000	15	100,000	15	100,000	15	100,000	15	100,000	15.0
E-1	75,000	..	35,700	..	22,700	..	16,300	..	12,500	..	10,000	..
Ex-1	25,000	60	64,300	23	77,300	19	83,700	18	87,500	17	90,000	16.5
E-2		..	39,300	..	25,000	..	18,000	..	13,750	..	11,000	..
Ex-2		..	25,000	60	52,300	29	65,700	23	73,750	20	79,000	19.0
E-3		..		..	27,300	..	19,500	..	15,000	..	12,000	..
Ex-3		..		..	25,000	60	46,200	33	58,750	25	67,000	22.0
E-4		..		..		..	21,200	..	16,250	..	13,000	..
Ex-4		..		..		..	25,000	60	42,500	35	54,000	28.0
E-5		..		..		..		..	17,500	..	14,000	..
Ex-5		..		..		..		..	25,000	60	40,000	37.5
E-6		..		..		..		..		..	15,000	..
Ex-6		..		..		..		..		..	25,000	60.0

NOTE: The data in the above table are not final, but only a rough approximation for the specific example discussed in the text and calculated for the purpose of facilitating other intermediate steps in determining the operating conditions.

In the first place, the total evaporation required will be 75 per cent. by weight $\frac{60 - 15 = 45}{60} = 0.75 = 75 \text{ per cent. or } 75,000 \text{ lbs. per hour.}$

Assume for the purpose of preliminary calculations that the evaporation in each body of a multiple effect, beginning with the first body will vary in the following proportions: 1.0, 1.1, 1.2, 1.3, 1.4, 1.5 The accompanying log above indicates the expected density in each body of the various multiple effects, and also the evaporation in each body under this convention.

The rise in the boiling point is shown below for each body condition, and also the total temperature losses from this source for each set.

Bodies	Single		Double		Triple		Quad.		Quint.		Sext.	
	Dens.	Loss	Dens.	Loss	Dens.	Loss	Dens.	Loss	Dens.	Loss	Dens.	Loss
First	60.0	9.5	23.0	2.5	19.0	2.0	18.0	1.5	17.0	1.5	16.5	1.5
Second	...	...	60.0	9.5	29.0	3.5	23.0	2.5	20.0	2.0	19.0	2.0
Third	...	...	...	...	60.0	9.5	33.0	4.0	25.0	3.0	22.0	2.5
Fourth	...	...	...	...	...	...	60.0	9.5	35.0	4.5	28.0	3.5
Fifth	...	...	...	...	...	...	...	...	60.0	9.5	37.5	5.0
Sixth	...	...	...	...	...	...	...	...	...	...	60.0	9.5
Total	9.5		12.0		15.0		17.5		20.5		24.0	

The temperature corresponding to 8 lbs. gauge is 235° F. , and that corresponding to 26 in. vacuum is 125° F. , so that the total drop is 110° F. There would then be $110 - 9.5 = 100.5^{\circ}$ drop for the single effect; $110 - 12 = 98^{\circ}$ for the double effect; $110 - 15 = 95^{\circ}$ for the triple effect; $110 - 17.5 = 92.5^{\circ}$ for the quadruple effect; $110 - 20.5 = 98.5^{\circ}$ for the quintuple effect; and $110 - 24 = 86^{\circ}$ for the sextuple effect.

It is possible and practicable to work any of these combinations, and as a matter of fact, they have all been worked except the single effect, which no one would consider to-day.

The method of procedure to arrive at the proper distribution of working temperature should be about as follows:

The discussion will be limited to one of these combinations, the case of the sextuple effect, which is the most complicated. There is a net working drop of 86° after deducting the boiling temperature losses. Corrections must be made for a further determinable loss which can be estimated with a fair degree of accuracy, that due to hydrostatic head. (See Chapter 5.) By making an approximate estimate of the probable pressures in the various bodies, these would be about as follows: 1.0° , 1.5° , 2.0° , 2.5° , 4.0° , 9.0° , total, 20° , which, when deducted from the available 86° leaves 66° net drop.

This drop must be divided among the various bodies in proportion to the amount of evaporation each is to do. It must also be distributed in the inverse proportion to the probable coefficient of heat transmission as affected by all other causes, such as viscosity, fouling, air binding, steam temperature, etc., which require careful scrutiny.

It is fair to assume for the given example the coefficients as outlined below:

Bodies	Expected Coef.	Expected Evap.	Drop Factor
1	750	10,000	13.3
2	700	11,000	15.7
3	650	12,000	18.5
4	550	13,000	23.6
5	450	14,000	31.1
6	225	15,000	66.7
Total			168.9

The total temperature drop of 66° must be divided up among the various bodies in proportion to the drop factor which is the quotient of the expected evaporation divided by the expected coefficient. This is shown in the following:

For No. 1 body, $66 \times \frac{13.3}{168.9} = 5.21$, approximately, 5.0 degrees			
For No. 2 body, $66 \times \frac{15.7}{168.9} = 6.13$,	"	6.0	"
For No. 3 body, $66 \times \frac{18.5}{168.9} = 7.23$,	"	7.5	"
For No. 4 body, $66 \times \frac{23.6}{168.9} = 9.22$,	"	9.0	"
For No. 5 body, $66 \times \frac{31.1}{168.9} = 12.13$,	"	12.5	"
For No. 6 body, $66 \times \frac{66.7}{168.9} = 26.10$,	"	26.0	"

The final total temperature distribution would be in accordance with the following table, where are also shown the corresponding vacua and pressures:

DISTRIBUTION OF TEMPERATURES, PRESSURES, AND LATENT HEAT.

Bodies	S. B. I	1	2	3	4	5	6
B. T. loss.....		1.5	2.0	2.5	3.5	5.0	9.0
S. H. loss.....		1.0	1.5	2.0	2.5	4.0	9.0
Net drop		5.0	6.0	7.5	9.0	12.5	26.0
Total drop		7.5	9.5	11.5	15.0	21.5	44.0
Vapor temperature	235.0	226.5	217.0	205.5	190.5	169.0	125.0
Latent heat	955.4	961.0	967.2	974.4	983.6	996.4	1,021.6
Liquid temperature ...		228.0	219.0	208.0	194.0	174.0	134.0
P. and vac.....	8.00#	5.00#	1.50#	4" V.	11" V.	18" V.	26" V.

It is now possible to make the heat balance for the problem in question. The condensate from the steam belt of the first effect will be removed separately. That from the others will be passed from one body to the

next giving up its sensible heat at each drop, and the sum total of all these condensates will be removed from the steam belt of the last effect by means of a suitable pump.

With proper heat insulation, the radiation losses will be practically negligible, and will therefore be neglected.

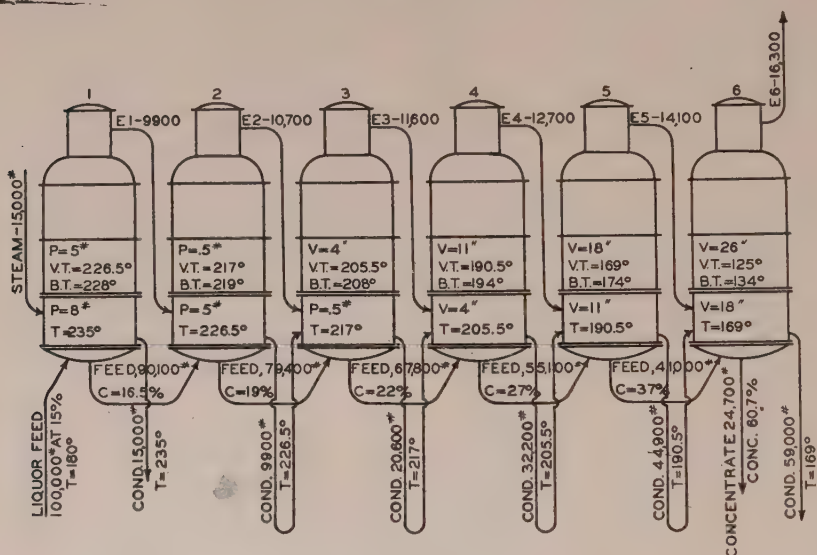


FIG. 3.—Diagrammatic Sketch of Sextuple Effect. See Heat Balance.

Fig. 3, a diagrammatic sketch, shows the proposed cycle with the flow of vapor, liquor, and condensate indicated as well as the temperatures and densities at all points. This data is taken from the heat balance and the temperature distribution, table above, and shows practically all the conditions worthy of consideration in the design and operation for the work contemplated.

HEAT BALANCE SEXTUPLE EFFECT EVAPORATOR.

(With By-passed Condensate.)

Bodies	Specifications	Heat	Liquor
(1)	Steam, 15,000 lbs. at 235 ($L = 955.4$) = $15,000 \times 955.4 =$	14,320,000	100,000
	Deduct for heating, $100,000 \times (228 - 180 = 48) = \dots$	4,800,000	
	Available for evaporation.....	9,520,000	
	L at 226.5 = 961. $E_1 = \frac{9,520,000}{961} = \dots$		9,900
	Transferred to No. 2 body.....		90,100

HEAT BALANCE SEXTUPLE EFFECT EVAPORATOR.

(With By-passed Condensate.)

Bodies	Specifications	Heat	Liquor
(2)	Vapor ex. No. 1 body.....	9,520,000	
	Liquor flash = $90,100 \times (228 - 219 = 9) =$	810,900	
	Available for evaporation.....	10,330,900	
	L at 217 = 967.2. $E_2 = \frac{10,330,900}{967.2} =$		10,700
	Transferred to No. 3 body.....		79,400
(3)	Vapor ex. No. 2 body.....	10,330,900	
	Liquor flash = $79,400 \times (219 - 208 = 11) =$	873,400	
	Condensate flash, $9,900 \times (226.5 - 217 = 9.5) =$	94,000	
	Available for evaporation.....	11,298,300	
	L at 205.5 = 974.4. $E_3 = \frac{11,298,300}{974.4} =$		11,600
	Transferred to No. 4 body.....		67,800
(4)	Vapor ex. No. 3 body.....	11,298,300	
	Liquor flash, $67,800 \times (208 - 194 = 14) =$	950,000	
	Condensate flash, $9,900 (E_1)$ $10,700 (E_2)$		
	$20,600 \times (217 - 205.5 = 11.5) =$	237,000	
	Available for evaporation.....	12,485,300	
	L at 190.5 = 983.6. $E_4 = \frac{12,485,300}{983.6} =$		12,700
	Transferred to No. 5 body.....		55,100
(5)	Vapor ex. No. 4 body.....	12,485,300	
	Liquor flash = $55,100 \times (194 - 174 = 20) =$	1,102,000	
	Condensate flash, $9,900 (E_1)$ $10,700 (E_2)$ $11,600 (E_3)$		
	$32,200 \times (205.5 - 190.5 = 15) =$	483,000	
	Available for evaporation.....	14,070,300	
	L at 169 = 996.4. $E_5 = \frac{14,070,300}{996.4} =$		14,100
	Transferred to No. 6 body.....		41,000
(6)	Vapor ex. No. 5 body.....	14,070,300	
	Liquor flash = $41,000 \times (174 - 134 = 40) =$	1,640,000	
	Condensate flash, $9,900 (E_1)$ $10,700 (E_2)$ $11,600 (E_3)$ $12,700 (E_4)$		
	$44,900 \times (190.5 - 169 = 21.5) =$	967,000	
	Available for evaporation.....	16,677,300	
	L at 125 = 1,021.6. $E_6 = \frac{16,677,300}{1,021.6} =$		16,300
	Concentrate ex No. 6.....		24,700

HEAT BALANCE SEXTUPLE EFFECT EVAPORATOR.
(Without By-passed Condensate.)

Bodies	Specifications	Heat	Liquor
(1)	Steam, 15,000 lbs. at 235 ($L = 955.4$) = $15,000 \times 955.4 =$	14,320,000	100,000
	Deduct for heating, $100,000 \times (228 - 180 = 48) = \dots$	4,800,000	
	Available for evaporation.....	9,520,000	
	L at 226.5 = 951. $E_1 = \frac{9,520,000}{951} = \dots$		9,900
	Transferred to No. 2 body.....		90,100
(2)	Vapor ex. No. 1 body.....	9,520,000	
	Liquor flash, $90,100 \times (228 - 219 = 9) = \dots$	810,000	
	Available for evaporation.....	10,330,000	
	L at 217 = 967.2. $E_2 = \frac{10,330,000}{967.2} = \dots$		10,700
	Transferred to No. 3 body.....		79,400
(3)	Vapor ex. No. 2 body.....	10,330,000	
	Liquor flash, $79,400 \times (219 - 208 = 11) = \dots$	873,400	
	Available for evaporation.....	11,203,400	
	L at 205.5 = 974.4. $E_3 = \frac{11,203,400}{974.4} = \dots$		11,500
	Transferred to No. 4 body.....		67,900
(4)	Vapor ex. No. 3 body.....	11,203,400	
	Liquor flash, $67,900 \times (208 - 194 = 14) = \dots$	951,000	
	Available for evaporation.....	12,154,400	
	L at 190.5 = 983.6. $E_4 = \frac{12,154,400}{983.6} = \dots$		12,360
	Transferred to No. 5 body.....		55,540
(5)	Vapor ex. No. 4 body.....	12,154,400	
	Liquor flash, $55,540 \times (194 - 174 = 20) = \dots$	1,110,800	
	Available for evaporation.....	13,265,200	
	L at 169 = 996.4. $E_5 = \frac{13,265,200}{996.4} = \dots$		13,300
	Transferred to No. 6 body.....		42,240
(6)	Vapor ex. No. 5 body.....	13,265,200	
	Liquor flash, $42,240 \times (174 - 134 = 40) = \dots$	1,689,600	
	Available for evaporation.....	14,954,800	
	L at 125 = 1,021.6. $E_6 = \frac{14,954,800}{1,021.6} = \dots$		14,640
	Concentrate ex. No. 6 body.....		27,600

Evaporation with Flash By-pass		Evaporation without Condensate By-pass	
E-1	9,900	E-1	9,900
E-2	10,700	E-2	10,700
E-3	11,600	E-3	11,500
E-4	12,700	E-4	12,360
E-5	14,100	E-5	13,300
E-6	16,300	E-6	14,640
E-T	75,300	E-T	72,400

Difference, $75,300 - 72,400 = 2,900$, or about 4 per cent.

Pounds Evaporation per Pound Steam,

$$\frac{75,300}{15,000} = 5.01$$

Evaporation per Pound Steam,

$$\frac{72,400}{15,000} = 4.83$$

In making up this heat balance, Marks and Davis steam tables have been used throughout. The same method has been used as illustrated in the preceding chapter with necessary modifications to take care of the different conditions.

In general, when calculating such balances, it is useless to waste time trying to make the figures check closer than a few hundred pounds, for variations of operating conditions will make a greater difference than such a slight inaccuracy, and the data assumed are subject to much variation in any case.

It is readily apparent that by following the general principles laid down in this discussion, it is possible to estimate correctly the work of evaporators with a fair degree of precision, and the data thus secured are invaluable to the designer and operator, for they enable him to base his work on figures about which there can be little dispute.

There are many variations of these heat balances as the cycles of operation are changed. This is especially true when vapor is withdrawn from the various bodies for heating purposes as shall be noted later.

As a matter of interest, the same heat balance has been calculated based on removal of condensate from each body independently without recovering the flash. There is a difference of about 4 per cent. in the work accomplished by the use of the same amount of steam, which is just one illustration of the light revealed by a careful heat balance analysis.

Under certain conditions it is necessary to make a correction for specific heat of this solution. This has not been done here, as it would not have materially affected the result, and it was unwise to complicate further the calculations which are already involved.

Chapter 9.

Vapor Heating.

The general designs and characteristics of solution heaters, as well as the heat transfers obtainable, have been discussed in another chapter under such a heading. The use of heaters in connection with multiple effects is so important, however, that it is treated separately here.

There is an excellent opportunity, not realized by the layman, to effect very material steam economies in plants equipped with evaporators where it is possible to combine the work of these with other necessary heating operations.

Heaters using either live or exhaust steam obtain only one stage of work from the heat supply, and this can be changed into multi-stage, conditions permitting, by the use of vapors from the various bodies of the evaporators, providing the temperatures of these vapors are sufficiently high to accomplish the work contemplated. With well-designed apparatus, it is usually possible to bring the temperature of liquids to be heated to within a few degrees of that of the steam or vapors used for this purpose. With good conditions liquor temperatures have been observed as close as within 2° F. of the vapor temperatures, although it is not safe to estimate on such high efficiencies. For the specific purpose of this discussion, it will be assumed that a difference of 8° can be maintained.

According as the temperature of the liquid to be heated is lower, so is it possible to use vapors from bodies nearer the condenser, which have already done useful work, and therefore are less expensive and valuable, with consequent greater economy. It may even be possible to do a substantial amount of heating with vapors leaving the evaporator on their way to the condenser, after the heat has done all the possible useful work, if the initial temperature of the liquor is relatively cold, in which case, the heater becomes in reality a surface condenser, taking off part of the load, the balance of the vapors going to the main condenser. The saving in B.t.u.'s in this case, as compared with live or exhaust heating is equivalent to the total amount of heat so utilized. As the vapors are taken from bodies further and further from the condenser, the saving is less and less in proportion.

Rillieux recognized from the beginning the economy of such combinations, and formulated what he called his *Second Principle*, which may

be stated as follows: If vapors from the various bodies of a multiple effect are utilized for heating operations instead of steam, the saving so accomplished is equal to the amount of vapors so utilized multiplied by a factor indicating the position of the cell in the set, beginning with the first effect, and divided by the total number of bodies in the series. Karl Abraham gives the following table representing this saving when 100 lbs. of vapor are taken for heating from the various bodies of an evaporator:

Bodies	1	2	3	4	5	6
Single effect	100.0					
Double effect	50.0	100.0				
Triple effect	33.3	66.7	100.0			
Quadruple effect	25.0	50.0	75.0	100.0		
Quintuple effect	20.0	40.0	60.0	80.0	100.0	
Sextuple effect	16.7	33.3	50.0	66.7	83.3	100.0

The savings as shown by this table are only approximately true, in view of the fact that they are not based on a true heat analysis, and they will be subject to more or less variations, as the conditions of operation vary, as has been pointed out in the discussion of basic principles.

Perhaps a better way to express the truth about savings by the use of vapor heaters would be as follows: The saving of heat by the use of vapors from bodies of a multiple effect for heating operations is equal to the total amount of latent heat in the vapors so condensed, less the cost of replacing the evaporation which would have been accomplished by their use in the evaporator, had they not been withdrawn.

It is a fact often overlooked, however, that when there are heating operations in a plant in which there is a multiple effect it is possible to secure very substantial steam economies by intelligently combining these two operations, the actual saving depending upon the range of temperatures, and the amount of heating necessary.

Thus, in the cane sugar factory, all of the juices come from the mills at temperatures varying from 70° F. to 100° F., depending upon the location of the factory, the process used, and climatic conditions. These juices must be heated to approximately 212° F. in the defecation, a range of from 112° F. to 142° F. It is further possible to split this heating into several stages to correspond to the probable heating that can be accomplished by vapors from the various bodies of the evaporating apparatus, beginning with vapors from the last body, and going counter current, finishing with vapors from the first effect. This naturally is an extreme case, but modifications of the cycle are in quite common use.

There are many applications in the various industries, and the scheme being correct in principle, there is nothing to prevent its use wherever it can be operated to advantage. The objection is that it ties up the heating and evaporating operations somewhat too intimately. This can be easily overcome by arranging valved connections so that vapors can

be cut out and steam substituted whenever it becomes advisable or necessary. The returns from such an arrangement are well worth the trouble.

In view of the importance of the above considerations, there is submitted below a complete analysis of the possible combinations of heating

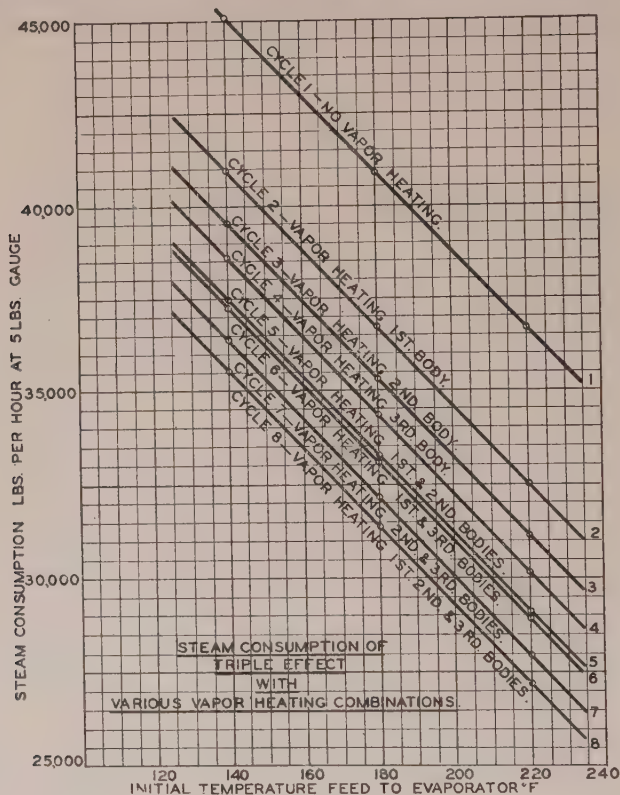


Fig. 9.

and evaporation for a triple effect which is to operate under assumed conditions to accomplish the following work as outlined:

1. Heat 100,000 lbs. per hour of a given solution from 80° F. to 212° F.
2. After other intermediate processes are completed, concentrate this solution from its initial density of 15 per cent. solids to a final density of 60 per cent. solids, thus evaporating 75,000 lbs. of water per hour. The temperature of the solution as it enters the evaporator is taken to be 180° F.
3. The conditions of operation are arbitrarily assumed as follows:

Available steam pressure.....	5 lbs. gauge
Vacuum in first effect.....	1 in. hg.
Vapor temperature corresponding.....	210° F.
Boiling temperature first effect.....	211° F.
Vacuum in second effect.....	12 in. hg.
Vapor temperature corresponding.....	187° F.
Boiling temperature second effect.....	190° F.
Vacuum third effect.....	24 in. hg.
Vapor temperature corresponding.....	141° F.
Boiling temperature third effect.....	150° F.

It is also assumed that it will be possible by the use of proper vapor heaters to bring the temperature of this liquor up to within 8° F. of the vapor used in the heater.

There are eight combinations possible in the heating and evaporation of this liquor under prescribed conditions, which, for convenience and reference, are identified as follows:

Cycle 1. Evaporation triple effect, no vapor heating;

Cycle 2. Evaporation triple effect, vapor heating first effect supplemented with exhaust steam;

Cycle 3. Evaporation triple effect, vapor heating second effect supplemented with exhaust steam;

Cycle 4. Evaporation triple effect, vapor heating third effect supplemented with exhaust steam;

Cycle 5. Evaporation triple effect, vapor heating first and second effects, supplemented with exhaust steam;

Cycle 6. Evaporation triple effect, vapor heating first and third effects, supplemented with exhaust steam;

Cycle 7. Evaporation triple effect, vapor heating second and third effects, supplemented with exhaust steam;

Cycle 8. Evaporation triple effect, vapor heating first, second, and third effects, supplemented with exhaust steam.

Heat balances have been calculated, and are herewith submitted showing complete heat analyses for these various *cycles*, and abstracts of each are shown in the illustrations accompanying, which indicate the flow of steam, vapor, and liquor throughout the equipment, as well as temperatures, pressures, vacua existing at all points.

A summary log is tabulated, showing the useful information thus developed, and Fig. 9 shows graphically a comparison of the steam requirements for the various cycles, indicating as well the effect of changing the initial temperature of the liquor as it enters the apparatus.

SUMMARY LOG.

Cycle	Evap.-1	Evap.-2	Evap.-3	Total Evap.	St. to Evap.	St. to Heat	Total Steam
1.....	23,700	25,000	26,300	75,000	27,200	13,750	40,950
2.....	32,150	20,800	22,050	75,000	35,750	1,040	36,790
3.....	27,200	28,300	19,500	75,000	30,700	3,430	34,430
4.....	23,700	25,000	26,300	75,000	27,200	8,220	35,400
5.....	28,750	27,500	18,700	74,950	32,300	1,040	33,340
6.....	28,450	22,600	23,900	74,950	32,100	1,040	33,140
7.....	25,300	26,500	23,200	75,000	28,800	3,430	32,230
8.....	26,900	25,700	22,400	75,000	30,400	1,040	31,440

SUMMARY Log (Continued).

Cycle	Cond. Vap. H. 1 Eff.	Cond. Vap. H. 2 Eff.	Cond. Vap. H. 3 Eff.	Cond. Load	Actual Economy	Rillieux Economy
1.....				26,300		
2.....	12,550			22,050	4,160	4,180
3.....		10,040		19,500	6,520	6,700
4.....			5,250	21,050	5,550	5,250
5.....	2,365	10,040		18,700	7,610	7,480
6.....	7,100		5,250	18,650	7,810	7,617
7.....		4,660	5,250	17,950	8,720	8,360
8.....	2,365	4,660	5,250	17,150	9,510	9,140

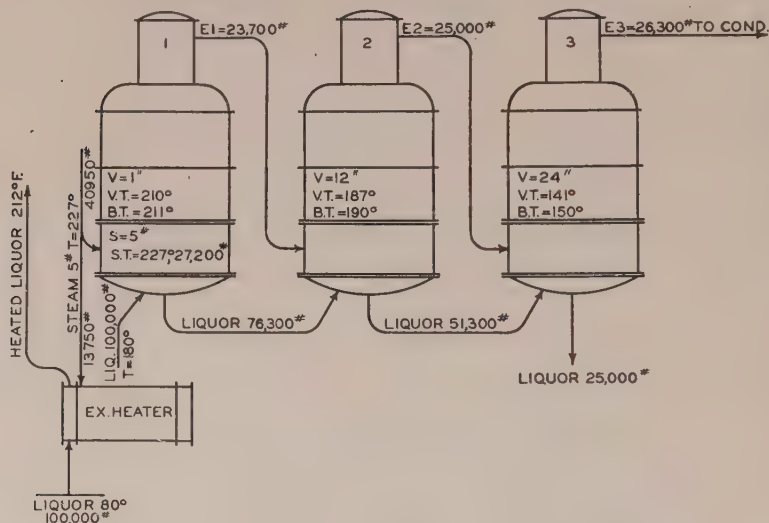


FIG. 1.—Cycle I—Triple—No Vapor Heating.

HEAT BALANCE CYCLE I.

NO VAPOR HEATING.

Bodies	Specifications	Heat	Liquor
(1) Steam, 27,200 lbs. at 227 = $27,200 \times (L = 960.7) =$		26,150,000	100,000
Deduct for heating, $100,000 \times (211 - 180 = 31) =$		3,100,000	
Available for evaporation.....		23,050,000	
L at 210 = 971.6. $E_1 = \frac{23,050,000}{971.6} =$			23,700
Transferred to No. 2.....			76,300
(2) Vapor ex. No. 1.....		23,050,000	
Liquor flash, $76,300 \times (211 - 190 = 21) =$		1,600,000	
Available for evaporation.....		24,650,000	
L at 187 = 985.7. $E_2 = \frac{24,650,000}{985.7} =$			25,000
Transferred to No. 3.....			51,300

HEAT BALANCE CYCLE I.

NO VAPOR HEATING.

Bodies	Specifications	Heat	Liquor
(3) Vapor ex. No. 2.....		24,650,000	
Liquor flash, $51,300 \times (190 - 150 = 40) =$		2,052,000	
Available for evaporation.....		26,702,000	
L at $141 = 1,012.6$. $E_3 = \frac{26,702,000}{1,012.6} =$			26,300
Concentrated liquor ex. No. 3.....			25,000
Requirements for heater $= \frac{100,000 \times (212 - 80 = 132)}{L \text{ at } 227 = 960.7} =$			13,750

If initial temperature of liquor to evaporator is increased or decreased 40° , the steam consumption will be increased or decreased by $\frac{100,000 \times 40}{960.7} = 4170$ lbs. This is true for all cycles, and will not be repeated in other balances.

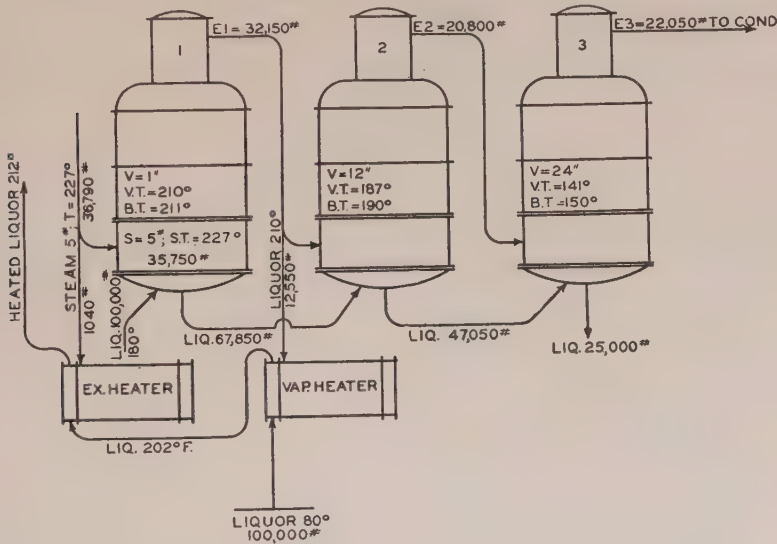


FIG. 2.—Cycle 2—Triple with Vapor Heating. First Effect Only.

HEAT BALANCE CYCLE 2.

VAPOR HEATING FIRST EFFECT.

Bodies	Specifications	Heat	Liquor
(1) Steam, 35,750 lbs. at $227 = 35,750 \times (L = 960.7)$		34,350,000	100,000
Deduct for heating, $100,000 \times (211 - 180 = 31)$		3,100,000	
Available for evaporation.....		31,250,000	
L at $210 = 971.6$. $E_1 = \frac{31,250,000}{971.6} =$			32,150
Transferred to No. 2.....			67,850

HEAT BALANCE CYCLE 2.

VAPOR HEATING FIRST EFFECT.

Bodies	Specifications	Heat	Liquor
(2) Vapor ex. No. 1.....		31,250,000	
Liquor flash, $67,850 \times (211 - 190 = 21) =$		1,425,000	
Available heat		32,675,000	
Ded. for V. H., $100,000 \times (202 - 80 = 122) =$		12,200,000	
Available for evaporation.....		20,475,000	
L at $187 = 985.7$. $E_s = \frac{20,475,000}{985.7} =$			20,800
Transferred to No. 3.....			47,050
(3) Vapor ex. No. 2.....		20,475,000	
Liquor flash, $47,050 \times (190 - 150 = 40) =$		1,882,000	
Available for evaporation.....		22,357,000	
L at $141 = 1,012.6$. $E_s = \frac{22,357,000}{1,012.6} =$			22,050
Concentrated liquor ex. No. 3.....			25,000
Requirements for heater, $\frac{100,000 \times (212 - 202 = 10)}{960.7} =$			1040

HEAT BALANCE CYCLE 3.

VAPOR HEATING SECOND EFFECT.

Bodies	Specifications	Heat	Liquor
(1) Steam, 30,700 lbs. at $227 = 30,700 \times (L = 960.7) =$..		29,500,000	100,000
Deduct for heating, $100,000 \times (211 - 180 = 31)$		3,100,000	
Available for evaporation.....		26,400,000	
L at $210 = 971.6$. $E_s = \frac{26,400,000}{971.6} =$			27,200
Transferred to No. 2.....			72,800
(2) Vapor ex. No. 1.....		26,400,000	
Liquor flash, $72,800 \times (211 - 190 = 21) =$		1,525,000	
Available for evaporation		27,925,000	
L at $187 = 985.7$. $E_s = \frac{27,925,000}{985.7} =$			28,300
Transferred to No. 3.....			44,500
(3) Vapor ex. No. 2.....		27,925,000	
Liquor flash, $44,500 \times (190 - 150 = 40) =$		1,780,000	
Available heat		29,705,000	
Ded. for heater, $100,000 \times (179 - 80 = 99) =$		9,900,000	
Available for evaporation.....		19,805,000	
L at $141 = 1,012.6$. $E_s = \frac{19,805,000}{1,012.6} =$			19,500
Concentrated liquor ex. No. 3.....			25,000
Requirements for heater, $\frac{100,000 \times (212 - 179 = 33)}{960.7} =$			3430

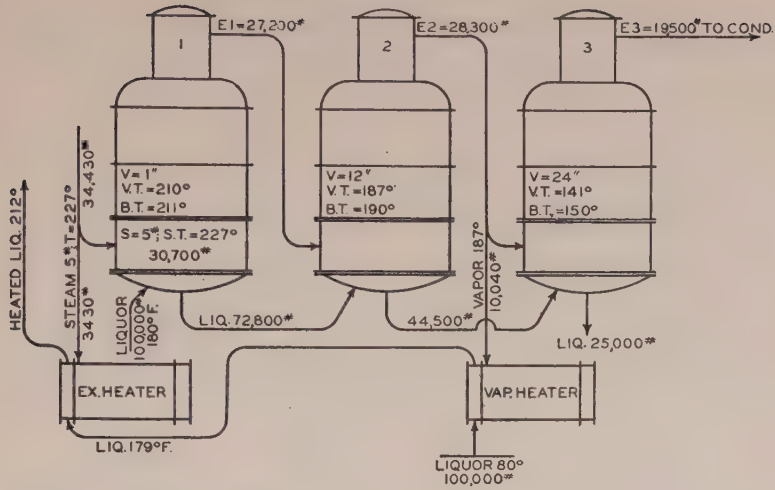


FIG. 3.—Cycle 3—Triple with Vapor Heating. Second Effect Only.

HEAT BALANCE CYCLE 4.

VAPOR HEATING THIRD EFFECT.

The heat balance for this cycle is exactly the same as for cycle 1, and will therefore not be repeated here.

The amount of exhaust steam required by the heater, however, is quite different, as the range of heating in this case is from 133° to 212°, instead of from 80° to 212° in the case of cycle 1. The steam requirements of this heater are therefore:

$$\frac{100,000 \times (212 - 133 = 79)}{960.7} = 8220$$

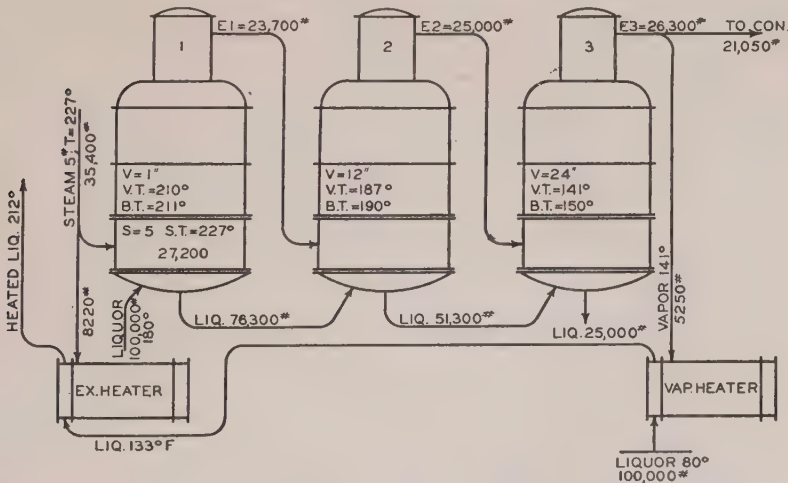


FIG. 4.—Cycle 4—Triple with Vapor Heating. Third Effect Only.

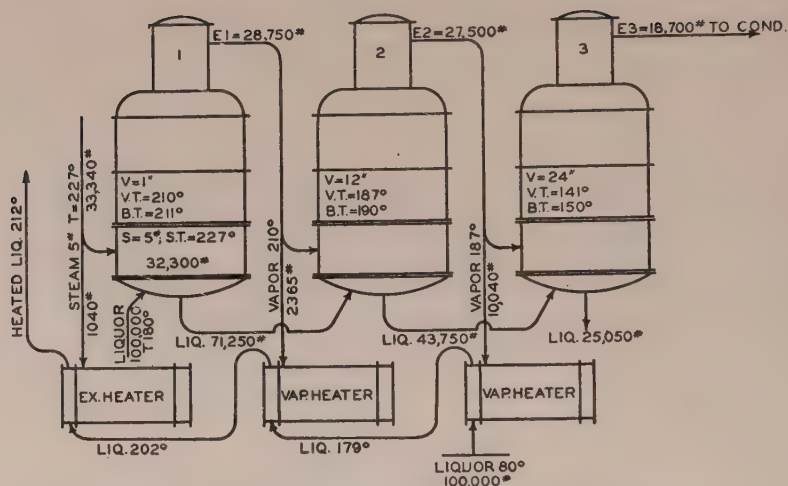


FIG. 5.—Cycle 5—Triple with Vapor Heating. First and Second Effects.

HEAT BALANCE CYCLE 5.

VAPOR HEATING FIRST AND SECOND EFFECTS.

Bodies	Specifications	Heat	Liquor
(1) Steam, 32,300 lbs. at 227° =	$32,300 \times (L = 760.7) = ..$	31,000,000	100,000
Deduct for heating, 100,000 × (211° — 180° = 31) = ..		3,100,000	
Available for evaporation.....		27,900,000	
L at 210° = 971.6. $E_1 = \frac{27,900,000}{971.6} = ..$			28,750
Transferred to No. 2.....			71,250
(2) Vapor ex. No. 1.....		27,900,000	
Add liquor flash, 71,250 × (211° — 190° = 21) = ..		1,495,000	
Available heat		29,395,000	
Deduct for heating, 100,000 × (202° — 179° = 23) = ..		2,300,000	
Available for evaporation		27,095,000	
L at 187° = 985.7. $E_2 = \frac{27,095,000}{985.7} = ..$			27,500
Transferred to No. 3.....			43,750
(3) Vapor ex. No. 2.....		27,095,000	
Add liquor flash, 43,750 × (190° — 150° = 40) = ..		1,750,000	
Available heat		28,845,000	
Deduct for heating, 100,000 × (179° — 80° = 99) = ..		9,900,000	
Available for evaporation		18,945,000	
L at 141° = 1,012.6. $E_3 = \frac{18,945,000}{1,012.6} = ..$			18,700
Concentrated liquor ex. No. 3.....			25,050
Heater requirements, $\frac{100,000 \times (212° - 202° = 10)}{1,012.6} = ..$			1,040

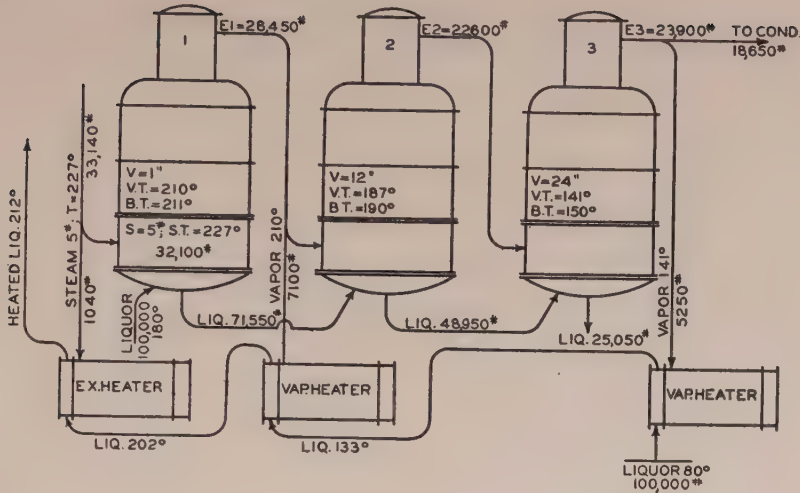


FIG. 6.—Cycle 6—Triple with Vapor Heating. First and Third Effects.

HEAT BALANCE CYCLE 6.

VAPOR HEATING FIRST AND THIRD EFFECTS.

Bodies	Specifications	Heat	Liquor
(1) Steam, 32,100 lbs. at 227 = 32,100 × (L = 960.7) = ..		30,750,000	100,000
Deduct for heating, 100,000 × (211 — 180 = 31)		3,100,000	
Available for evaporation		27,650,000	
L at 210 = 971.6. $E_1 = \frac{27,650,000}{971.6} =$			28,450
Transferred to No. 2			71,550
(2) Vapor ex. No. 1		27,650,000	
Liquor flash, 71,550 × (211 — 190 = 21) =		1,500,000	
Available heat		29,150,000	
Deduct for heating, 100,000 × (202 — 133 = 69) = ..		6,900,000	
Available for evaporation		22,250,000	
L at 187 = 985.7. $E_2 = \frac{22,250,000}{985.7} =$			22,600
Transferred to No. 3			48,950
(3) Vapor ex. No. 2		22,250,000	
Liquor flash, 48,950 × (190 — 150 = 40) =		1,958,000	
Available for evaporation		24,208,000	
L at 141 = 1,012.6. $E_3 = \frac{24,208,000}{1012.6} =$			23,900
Concentrated liquor ex. No. 3			25,050
Heater requirements, $\frac{100,000 \times (212 - 202 = 10)}{960.7} =$..			1040

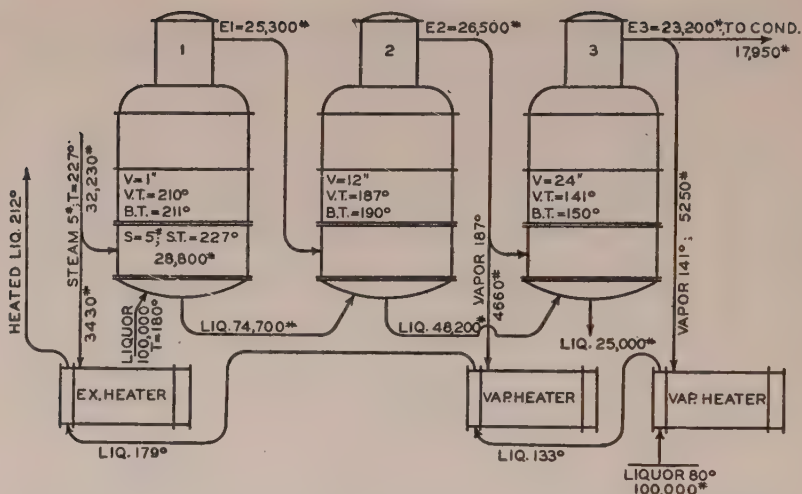


FIG. 7.—Cycle 7—Triple with Vapor Heating. Second and Third Effects.

HEAT BALANCE CYCLE 7.

VAPOR HEATING SECOND AND THIRD EFFECTS.

Bodies	Specifications	Heat	Liquor
(1) Steam, 28,800 lbs. at 227° = 28,800 × (L = 960.7) = ..		27,650,000	100,000
Deduct for heating, 100,000 × (211 — 180 = 31)		3,100,000	
Available for evaporation		24,550,000	
L at 210 = 971.6. $E_1 = \frac{24,550,000}{971.6} =$			25,300
Transferred to No. 2			74,700
(2) Vapor ex. No. 1		24,550,000	
Liquor flash, 74,700 × (211 — 190 = 21) =		1,568,000	
Available for evaporation		26,118,000	
L at 187 = 985.7. $E_2 = \frac{26,118,000}{985.7} =$			26,500
Transferred to No. 3			48,200
(3) Vapor ex. No. 2		26,118,000	
Liquor flash, 48,200 × (190 — 150 = 40) =		1,928,000	
Available heat		28,046,000	
Deducted for heating, 100,000 × (179 — 133 = 46) = ..		4,600,000	
Available for evaporation		23,446,000	
L at 141 = 1,012.6. $E_3 = \frac{23,446,000}{1,012.6} =$			23,200
Concentrated liquor ex. No. 3			25,000
Heater requirements, $\frac{100,000 \times (212 - 179 = 33)}{960.7} =$..			3430

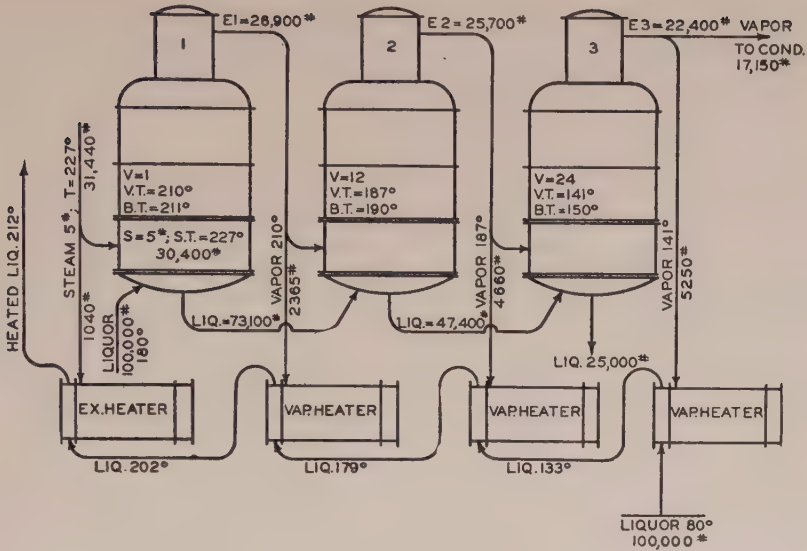


FIG. 8.—Cycle 8—Triple with Vapor Heating. All Bodies.

HEAT BALANCE CYCLE 8.

VAPOR HEATING, FIRST, SECOND AND THIRD EFFECTS.

Bodies	Specifications	Heat	Liquor
(1) Steam, 30,400 lbs. at 227°	$30,400 \times (L = 960.7) = ..$	29,200,000	100,000
Deduct for heating, 100,000 × (211 — 180 = 31) =...		3,100,000	
Available for evaporation.....		26,100,000	
L at 210 = 971.6. $E_1 = \frac{26,100,000}{971.6} = ..$			26,900
Transferred to No. 1.....			73,100
(2) Vapor ex. No. 1.....		26,100,000	
Liquor flash, 73,100 × (211 — 190 = 21) =.....		1,530,000	
Available heat		27,630,000	
Deduct for heating, 100,000 × (202 — 179 = 23) =....		2,300,000	
Available for evaporation.....		25,330,000	
L at 187 = 985.7. $E_2 = \frac{25,330,000}{985.7} = ..$			25,700
Transferred to No. 3.....			47,400
(3) Vapor ex. No. 2.....		25,330,000	
Liquor flash, 47,400 × (190 — 150 = 40) =		1,896,000	
Available heat		27,226,000	
Deduct heating, 100,000 × (179 — 133 = 46) =.....		4,600,000	
Available for evaporation.....		22,626,000	
L at 141 = 1,012.6. $E_3 = \frac{22,626,000}{1,012.6} = ..$			22,400
Concentrated liquor ex. No. 3.....			25,000
Heater requirements, $\frac{100,000 \times (212 - 202 = 10)}{960.7} = ..$			1040

The most interesting combination is cycle 8, and it is worthy of a few passing words. Attention has been called to the fact that the outstanding objection is the complication involved. It would seem, however, that the additional economies to be obtained in this instance justify the difficulties feared.

In actual operation, the variations will be felt most when cleaning individual heaters, a step to be expected at fairly regular intervals. This is not so serious as it appears at first glance, for the heaters are cleaned one at a time, under which conditions, cycle 8 is temporarily shifted to operate as cycle 5, 6, or 7, according to which heater is being cleaned. The resulting disturbance is not very serious, as can be seen by referring to the summary log. The maximum variation on the load to the condenser is only 1550 lbs., or about 9 per cent. increase. The evaporation in the various bodies would also change as shown in the log, the largest individual change being the evaporation in the last body which for cycle 5 is 18,700 lbs., as compared with 22,400 lbs. for cycle 8, a net difference of 3,700 lbs., which should be taken care of with comparative ease.

If the liquor being heated is shut off altogether, the operation automatically reverts to cycle 1, the biggest change involved being in the evaporation of the third effect, which increases from 22,400 to 26,300, or 3,900 lbs. This also can be accommodated with a little care.

The presumption is that the vacua and pressures will be as assumed, and in designing the apparatus, the proportions must be correct for such a performance. If not, the cycle will vary accordingly, but, unless there is a radical change, the net total results will not be affected greatly.

The economies effected by the use of the vapor heater on the third cell is a direct function of the difference between the initial temperature of the liquor to be heated and the temperature of the vapors from the last effect, and such economies will disappear as these two temperatures approach each other.

The method of calculation is general, and can be applied to a multiple effect having any number of bodies. It is somewhat laborious, but unquestionably represents the facts with dependable accuracy.

Chapter 10.

Vapor Compression.

It is stated in the literature on the subject that as early as 1857, a plan was proposed and tried whereby vapor leaving an evaporator was compressed by a steam injector for the purpose of reintroducing the mixture into the steam compartment. The results must have been quite unsatisfactory for it was not put into successful practice until within recent years. So far as the author knows, the first commercially successful installation of this type was made by Prache et Bouillon, of Paris, France.

Since then, Paul Kestner, also a Frenchman, and the designer of the well-known evaporator which bears his name, has also placed one on the market purporting to accomplish the results contemplated. There have also been some similar German and Swedish designs, some of which use turbocompressors instead of injectors, showing that the trend of thought is in that direction. There are also some recent American patents covering the same idea.

The possibilities of this cycle offer great promise, and the author believes that at some time in the future, very extensive and useful applications of vapor compression will be seen.

The theory is fairly simple, and very easily understood. The difficulty lies in the proverbial mechanical inefficiency of the injector, and also the lack of knowledge always incident to the solution of problems for which there are no good precedents. There are to-day steam jet vacuum pumps, which involve the same principles, and it should therefore be possible to devise an efficient steam injector capable of accomplishing the necessary work.

Refer to Fig. 1. Suppose the evaporator shown were to operate with its vapor space at atmospheric pressure, and that steam were maintained in the calandria at 3 lbs. gauge. It will be seen by referring to Marks and Davis steam tables that every pound of vapor leaving the apparatus contains 1150.4 B.t.u.'s, whereas every pound in the calandria contains at 3 lbs. gauge pressure, 1153.8 B.t.u.'s, the difference between the two being only 3.4 heat units, or $1/3$ per cent. The problem, therefore, consists of adding 3.4 B.t.u.'s to each pound of vapor leaving the evaporator to convert it back into steam available for reintroduction into the steam belt at 3 lbs. gauge pressure, which would seem to be a simple matter.

If a supply of steam at high pressure is available, it would seem easy to accomplish this result by mixing this high pressure steam with vapor from the evaporator at atmospheric pressure, giving a mixture at the right pressure. Thus, if live steam is available at 100 lbs. gauge, having

a total heat as shown in the steam tables of 1188.8 B.t.u.'s per pound, if this pressure is reduced to that of the calandria of the evaporator, 3 lbs., wherein the total heat is only 1153.8, there would be an excess of heat available equivalent to the difference between the two, or 35 B.t.u.'s, which could be utilized in supplying the deficiency of heat required to convert the 0 lbs. vapor into 3 lbs. vapor. Inasmuch as it only requires 3.4 B.t.u.'s per pound to do this, with each pound of live steam, it should be possible to boost up $\frac{35}{3.4} = 10.3$ lbs. of vapor. This means that the evaporator operating in this way would evaporate $10.3 + 1.0$ or 11.3 lbs. of water per pound of steam supplied. If the difference between the two sides of the

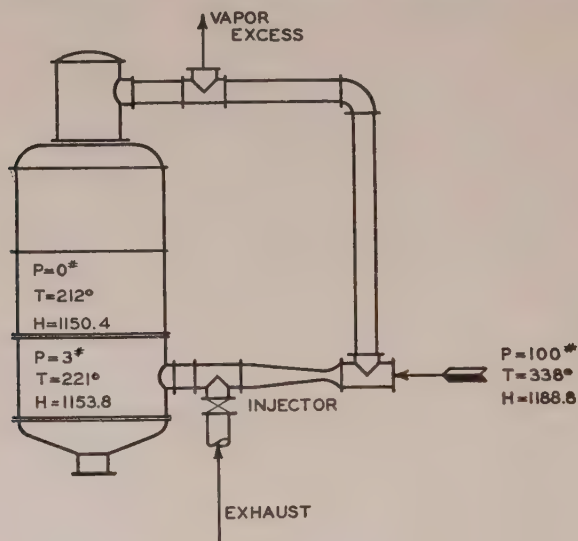


FIG. 1.

heating surface is increased, the difference between the total heats for the two pressures increases, and conversely. Therefore, a smaller operating drop across the heating surfaces means a greater ratio of evaporation per pound of live steam. The nearest approach to this theoretical work as per pressure conditions specified above according to the author's experience is about 3 lbs. of evaporation per pound of live steam used.

It would seem that the above data does not truly represent the steps necessary for the changes involved, and that it might be better to attack the problem from another angle. The steam injector problem suggested here, if the mixture is made by this means follows the laws of thermodynamics. It is a question of steam nozzles and velocities, expansions and compressions, and analysis by means of the Mollier Diagram might reveal information.

The expansions and compressions are substantially adiabatic, and

therefore, the entropy should remain constant. The amount of energy in the form of heat required to compress one pound of steam from 0 lbs. gauge to 3 lbs. gauge, or from 14.7 absolute to 17.7 absolute by reference to the Mollier diagram is 15 B.t.u.'s, and would be accompanied with a superheat of 30°. The live steam, expanding from 100 lbs. gauge, or 114.7 absolute, down to 14.7, the pressure at the end of the live steam nozzle, will yield 150.8 B.t.u.'s and will be wet, the quality being about 90 per cent.

The ratio of compression will therefore be the same by either method of calculation, but there are facts brought out by the latter method which are concealed in the former, the superheating of the compressed vapor balancing the moisture in the expanded live steam.

It is evident that the vapor to be compressed must be dry, otherwise, the moisture contained will re-evaporate through the injector, robbing some of the valuable margin of heat from the live steam.

The limits of vapor compression are well defined, and it is necessary to select conditions favorable to their application if good results are expected. In general, it will be true that vapor compression will yield satisfactory economies when the difference of temperature required to operate the evaporator is small, and such installations must therefore be limited to the concentration of low density, easy boiling solutions of low viscosity. It will also be true for the same reason, that better results will be obtained with high efficiency evaporators capable of doing substantial work with small temperature differences. On account of the large specific vapor volumes under vacuum, the probable successful range of operation will be at atmospheric pressure or above it.

The so-called thermocompressor should not be installed on evaporators where the difference of pressure between the two sides of the heating surface is subject to considerable fluctuations, as the nozzles are designed to operate under a given set of pressure conditions, and will become less efficient when these are changed. Under adverse conditions this thermocompressor will cease to work, and actually "back fire," when the live steam will flow back on itself, not having sufficient mass velocity to overcome the back pressure under the nozzle proportions.

Where the cost of power is low, it is possible to substitute a mechanical compressor for the steam injector. There are compressors of this sort in actual operation on evaporators doing this kind of work. They are very expensive in first cost, and therefore of limited application.

Chapter II.

Vapor Pipes.

The subject of vapor pipes and their proportions, usually neglected, is worthy of careful attention.

Hausbrandt, in his standard work on evaporation, has given tables in which he has worked out the friction losses in vapor pipes for various velocities and lengths, the data having been secured, no doubt, from actual observation, and subsequently interpolated for intermediate conditions.

It has been the experience of the authors that this is only part of the total pressure loss from the vapor side of one effect to the steam side of the other next in the series. There are at least two other factors which cannot be overlooked.

The most important of these is the loss in head, or difference of pressure necessary to bring the vapor velocity from the normal quiescence existing in the vapor space, up to the speed necessary to enter the catchall, and then to again accelerate this velocity to the speed in the piping leading from the catchall to the steam belt of the next effect.

The other source of loss is largely preventable, consisting of the friction or entrance head into the steam compartment itself.

As to the losses brought about by acceleration, it will be seen, by referring to Chapter 12, that in practically all the catchalls illustrated, there are involved at least two accelerations, and that, in addition for certain separators, such as the Hodek, there may be as much as six accelerations in series. This, however, is an extreme case, and in ordinary practice, it is safe to estimate on two accelerations, besides other points where there are swirls and eddies. These must be considered as part of the cost of operating and catchalls.

The ordinary friction such as is mentioned by Hausbrandt is in most cases smaller than one acceleration, and it has been the author's practice to estimate the total pressure drops from vapor belt to steam belt as equivalent to three accelerations, and this has been found to approximate the truth very closely.

In making the calculations, the formula $V = K\sqrt{2gh}$ is used very much in the same way as in estimating the liquor circulation. (See chapter on Natural Circulation.) V , represents the velocity of the vapors in feet per second; K is a constant whose value is assumed here as 0.63 due to the venal contraction of the stream; g is the value of acceleration due to gravity, 32.2; and H , is the head in feet. A specific example follows:

The vacuum in the third body of a quadruple effect is 18 in.; the

vapor speed is 100 ft. per second. What is the probable friction loss between the vapor belt of the third effect, and the steam belt of the fourth effect?

The known values in the equation are as follows:

$$\begin{aligned} V &= 100 \text{ ft. per sec.;} \\ K &= 0.63; \\ g &= 32.2. \end{aligned}$$

By substitution,

$$\begin{aligned} 100 &= 0.63 \sqrt{2 \times 32.2 \times H} = 63 \times 8.064 \times \sqrt{H} \\ \sqrt{H} &= \frac{100}{8.064 \times .63} = 19.65 \\ H &= (19.65)^2 = 386. \end{aligned}$$

In other words, in order to produce a speed in the vapor outlet of 100 feet per second, the column of vapor representing the head would have to be 386 feet high. Assume that the net cross section of this column were one square foot. The specific volume of steam at the assumed vacuum of 18 inches is 63.3 cubic feet per pound, and the weight of this column of vapor is therefore $\frac{386}{63.3} = 6.1$ lbs., which is the pressure per square foot for one acceleration, so that the pressure for three accelerations would be three times this amount, or 18.3 lbs. per sq. ft. This, when converted into inches of mercury is $\frac{18.3 \times 30}{144 \times 14.7} = 0.26$ inches.

On the basis of these calculations, the accompanying table and chart (Fig. 1) have been worked out. It is to be noted that these pressure losses are proportional to the square of the velocity, and also proportional directly to the density of the vapor, and in view of this fact, it is permissible to use higher velocities at high vacua, for the densities decrease enormously as the vacuum reaches above 24 inches, conditions usually met in the operation of the last body.

FRICITION LOSSES IN INCHES OF MERCURY FOR VARIOUS VACUA AND SPEEDS. See Fig. 1.

Vac.	651' Colmn.		1,164' Colmn.		2,610' Colmn.		4,650' Colmn.		7,260' Colmn.	
	Speed, 75'		Speed, 100'		Speed, 150'		Speed, 200'		Speed, 250'	
	Sp. V.	Wt. Col.	"Hg. Col.	Wt. Col.	"Hg. Col.	Wt. Col.	"Hg. Col.	Wt. Col.	"Hg. Col.	Wt. Col.
0.....	26.8	24.30	0.344	43.5	0.61	93.8	1.33	174.0	2.47	271.0
5.....	31.7	20.50	0.290	36.8	0.52	82.1	1.16	146.5	2.07	229.0
10.....	39.3	16.55	0.234	29.6	0.42	66.5	0.942	118.4	1.68	185.0
15.....	51.2	12.70	0.180	22.7	0.32	51.0	0.722	90.8	1.29	141.6
20.....	74.0	8.80	0.125	15.7	0.22	35.3	0.500	62.8	0.89	98.0
25.....	142.2	4.58	0.065	8.2	0.116	18.4	0.260	32.7	0.46	51.1
28.....	341.0	1.91	0.027	3.4	0.048	7.65	0.108	13.6	0.193	21.3

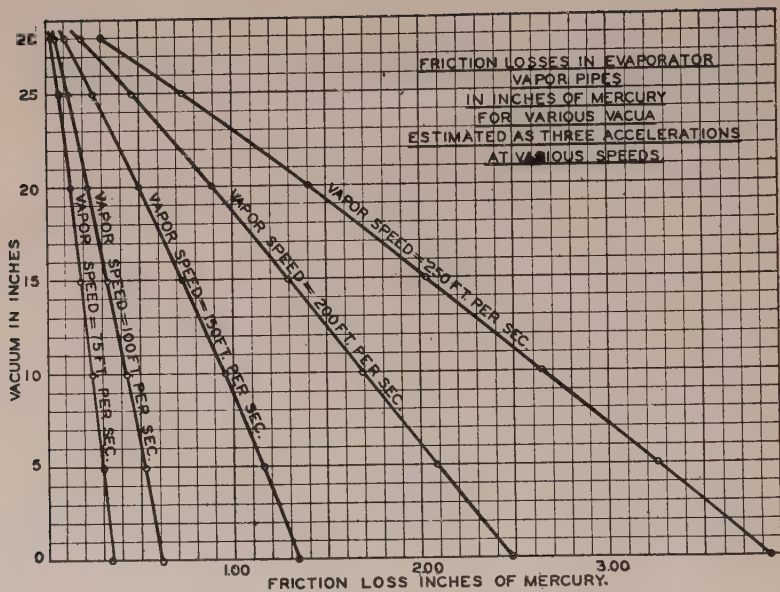


FIG. 1.

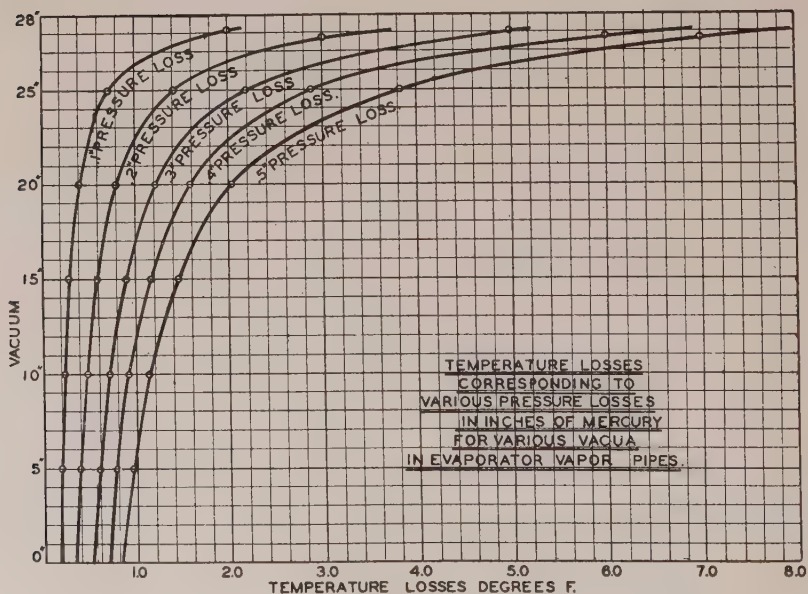


FIG. 2.

In appreciating the importance of these losses and their effect upon the transmission of heat in the evaporator, they should be converted into temperature losses, and thus their direct influence can be estimated. It is also well to bear in mind that inasmuch as the heat transfer at high temperature is much better than at low, a saving of temperature drop at low vacuum is much more beneficial than a corresponding saving at high vacuum as far as its net effect on the evaporator is concerned.

It is customary, in practice to use a speed of 100 feet per second in vapor pipes, except in the last effect, where for the reasons explained above, this is increased to 200 to 250 ft. per second. In turbine practice, this velocity is pushed up to 300 to 350 ft. per second, but it must be remembered that in this case, there is no catchall, and the vacua are higher as a rule. The vapor pipes moreover are short and direct, and therefore the friction losses smaller for given speeds.

Fig. 2 has been platted showing the temperature losses corresponding to given pressure losses when different vacua are maintained, thus a direct conversion can be made from the friction drops taken from the chart in Fig. 1. The range covers the usual conditions met in evaporator practice. The data for the chart is given in the log.

TEMPERATURE CORRESPONDING TO PRESSURE DROPS FOR
VARIOUS VACUA. See Fig. 2.

Vacuum	Loss for 0.1 in. Drop	Loss for 0.2 in. Drop	Loss for 0.3 in. Drop	Loss for 0.4 in. Drop	Loss for 0.5 in. Drop
28".....	1.80	3.45	5.00	6.40	7.80
25".....	0.70	1.40	2.20	2.90	3.60
20".....	0.40	0.80	1.20	1.60	2.00
15".....	0.30	0.60	0.90	1.15	1.45
10".....	0.23	0.46	0.70	0.90	1.13
5".....	0.20	0.40	0.60	0.80	1.00
0".....	0.15	0.30	0.45	0.60	0.75

As a matter of illustration, below is a table of friction and temperature losses in the vapor pipes of the sextuple effect which was analyzed in the chapter on the Practical Use of the Heat Balance. The vapor speeds, and vacua are shown in the table, and also the total overall temperature loss.

Index	Speed	Vac.	Pres. Loss Fig. 1	Temp. Loss Fig. 2
Vapor No. 1 to No. 2.....	100	0.5 lbs. p.	0.82 in.	1.10°
Vapor No. 2 to No. 3.....	100	0.5 in.	0.62 in.	1.00°
Vapor No. 3 to No. 4.....	100	4.0 in.	0.55 in.	1.04°
Vapor No. 4 to No. 5.....	100	11.0 in.	0.40 in.	0.95°
Vapor No. 5 to No. 6.....	100	18.0 in.	0.26 in.	0.95°
Vapor ex. No. 6.....	200	26.0 in.	0.36 in.	3.40°
			Total,	8.44°

Another factor which it is well to bear in mind is that the reduction of capacity involved by these temperature losses is not in proportion to the above as compared with the total temperature drop in the evaporator, but as compared with the *net* drop, which, in this particular instance is

66°, so that it is bad engineering to use vapor pipes which involve large temperature losses.

Fig. 3 shows a typical arrangement of vapor pipes and catchalls between two adjacent bodies, and the points are indicated at which the accelerations are known to take place.

With regard to the other point brought out at the beginning of the discussion, unless great care is used, the biggest pressure loss will occur as the steam enters the calandria. This is really due to improper design, and there is no good excuse for having serious losses at this point.

In cylindrical calandrias, the tubes immediately in front of the vapor inlet should be omitted, so as to permit free passage without constriction.

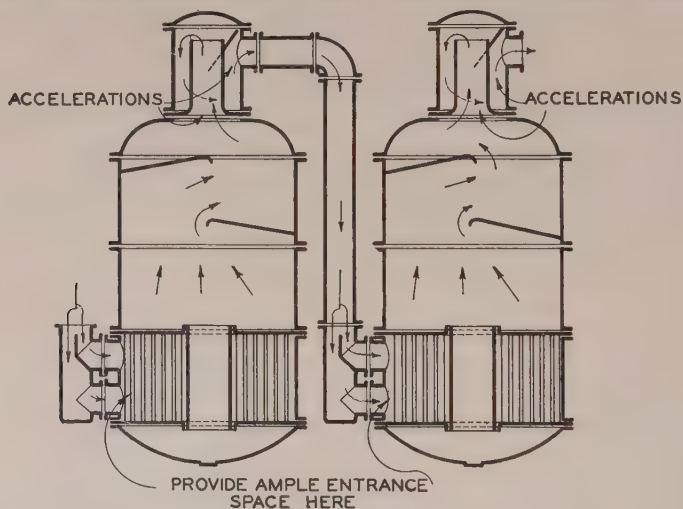


FIG. 3.—Typical Arrangement of Vapor Pipes and Catchalls Showing Important Points.

It has been the author's practice to use two circular inlets, one above the other as shown in the illustration, each of these openings having the same area as the vapor pipe, and to split the incoming vapor by means of a division plate so as to divide it equally between the two. Thus the entrance speeds are cut in two. In addition, there is a wedge-shaped opening left in the heating surface, obtained by leaving out tubes, thus permitting the free and uninterrupted flow of vapors. With these precautions, the losses of pressure at this point are almost negligible.

Where calandrias with annular spaces are used, unless the area in this passage is enlarged out of proportion, it is impossible to avoid material losses due to current interferences and sudden turns which must be taken at high speeds, besides poor distribution of steam to the heating surface. Attempts have been made to overcome the difficulty by using twin vapor pipes which produces the same results as if the cross section of the annular space had been doubled. Nevertheless, such losses are usually very considerable, especially in large evaporators.

Chapter 12.

Entrainment and its Prevention.

One of the most important details in connection with evaporator work is the problem of entrainment and the proper provisions to overcome it. This condition generally arises from one of three causes:

1) Foaming or frothing, which is mainly the result of evaporation of materials having low surface tension in unsuitable apparatus.

2) Carrying over, which is brought about by mechanical entrainment of small particles of solution by rapidly moving vapor currents.

3) Flashing, or sudden ebullition of the liquid mass due to sudden drop in temperature.

1) Taking up these topics in order, foaming, which results from low surface tension, is due to the occlusion of small volumes of vapor by a relatively strong film of liquid, which is hard to break, and in fact, which does not break except under special conditions. There is an accumulation of these bubbles superposed one above the other extending above the top tube sheet into the vapor space, very often filling the entire apparatus, and going over into the vapor pipe of the succeeding effect, or condenser, as the case may be, resulting in the loss of the material thus carried over.

Where a solution having this characteristic is to be evaporated, the logical procedure is to select an apparatus with long tubes, in which the velocity of circulation in the tubes is relatively high (over 12 ft. per second). In this case, it has been established that the froth is entirely broken up by the whipping effect of the rapidly moving streams of vapor and liquor leaving the upper part of the tubes. From the very nature of the case, such a procedure yields very fine droplets which are easily carried over by vapor currents, and therefore require very careful attention in the subsequent separation if losses are to be avoided.

It sometimes happens that a solution from which it is desired to crystallize out solids also possesses low surface tension. In this case, it is not permissible to use the long tube type evaporator, which would produce extremely fine crystals difficult to settle and remove from the system. It is then necessary to resort to some other means for overcoming this difficulty. The usual method is to introduce into the liquid in each body some foreign substance which will increase the surface tension. This trick is very old, and to the best of our knowledge, was used in the sugar industry on some of Rillieux's original apparatus. In fact, in the old days, all pans and evaporators were equipped with "butter cups," "grease cups," or "tallow cups." To this day, many manufacturers still provide

these accessories with their apparatus. They are almost never used, and are now being omitted except where specially needed.

Where such devices are employed, it is best to arrange them to feed continuously into the evaporator a small quantity of material, determined experimentally to be sufficient to overcome the trouble. The selection of this must be such as not to injure the solution. Many different materials have this property, such as butter, grease, tallow, soap, kerosene, etc. Some of these, such as kerosene, have the additional advantage of making the scale formation on the tubes less obstinate. The quantity required to accomplish the desired result is generally very small, as compared with the amount of solution flowing through the evaporator.

It is possible to overcome losses incident to frothing by the use of specially designed separators, in which an artificial whipping action is induced by the vapor currents, thus breaking up the bubbles, and permitting separation to take place effectively.

It is a fact worthy of note that whereas this foam is relatively light, nevertheless a deep accumulation of it has weight enough to cause appreciable loss of pressure between the upper tube sheet and the point of separation. This in turn, means a loss of temperature fall, and consequent impairment of capacity. This is more or less true as the character of the solution varies, but cases have been known where this loss was very considerable.

Foam has also been overcome by causing the froth to pass over heating surface which explodes the bubbles. This entails another difficulty in that such heating surface under the conditions tends to become coated rapidly with material actually dried and baked on it, due to the relatively small amount of liquid in the froth, necessitating a subsequent difficult cleaning of the heating surface, and also materially lowered heat transmission during actual operation.

Other means have also been used or attempted with generally poor success, such as mechanical agitators, steam jets, liquor jets, etc., the idea being to break up the foam by beating. Many fruitless attempts have demonstrated it is advisable to resort to these methods only when others have proved worthless.

2) Aside from foaming, very serious losses often occur from entrainment, or the mechanical carrying over of minute drops of liquid by vapor currents leaving the evaporator. Unless provisions have been made against it, this loss will be considerable, especially as the volume and speed of the vapors increase. Practically all manufacturers equip their evaporators with catchalls, or separators, to overcome this difficulty.

In the past, many have been partial to the short tube evaporators, because it was thought that by their use losses from entrainment would be held down to a minimum. With separators of questionable efficiency, and not properly installed, this was undoubtedly true. In fact, it has often been contended that it is impossible to separate fine particles of liquor from rapidly moving vapor. This the author denies but it is undoubtedly true that as vapor currents move faster and the entrained drops are smaller, their separation becomes more and more difficult.

One can illustrate this problem well in the following way. In a given tube, in which evaporation is taking place, all the vapor generated by its heating surface must leave through the upper end mixed with the circulating liquor part of which at least assumes the maximum speed of the vapor. The result is that neglecting other interferences, these drops left to themselves would be projected to a height corresponding to their velocity in accordance with the well known equation $V^2 = 2gH$; V representing the velocity in feet per second; H representing the height of projection in feet; and g representing the acceleration due to gravity, or 32.2. In a

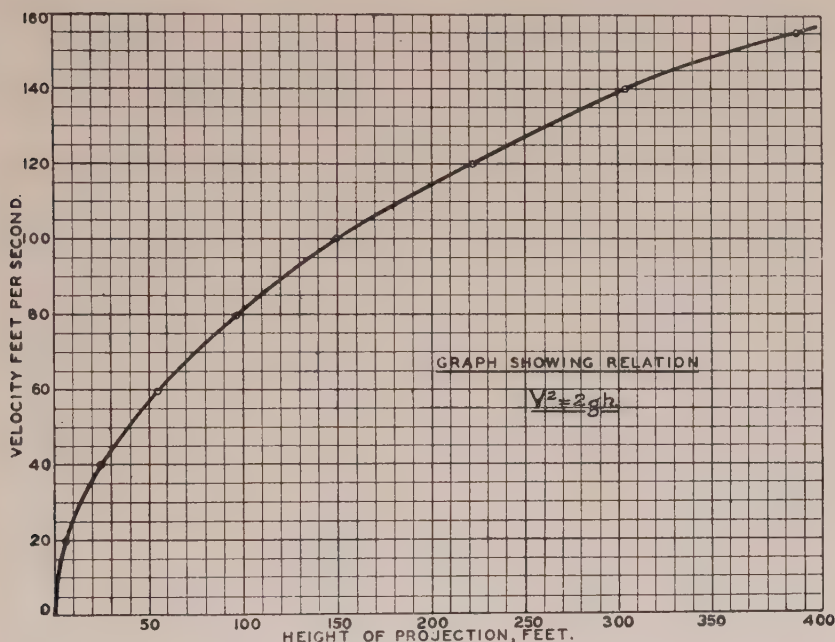


FIG. 1.

vertical tube evaporator the velocities vary according to vacuum conditions, proportions of tubes, rate of operation, from a minimum of about 2 ft. per second, to a maximum of about 125 ft. per second. Above is given a curve (Fig. 1) in which the value of H is shown for velocities within this range.

For long tube evaporators, much higher speeds prevail, sometimes attaining as high as 250 ft. per second, although such extremes are not advisable.

Inasmuch as the net vertical distance between the upper tube sheet of the evaporator and the dome or top is seldom more than 12 to 14 feet, it is evident from the curve that when the velocity at the end of the tubes

is from 20 to 25 feet per second, the initial speed of the drops of liquor leaving this point, if unaffected by other causes would project them up into the vapor outlet in the dome, beyond which the separator must be depended upon entirely to arrest them. This situation is further aggravated by the fact that the vapor currents are upwards, and that many of these drops are so small that they offer a very large area of contact to the currents as compared with their weights, and for that reason, it is a very easy matter for them to float along in the general direction of the vapor.

It is practically true that in almost every evaporator, the velocity of travel in the tubes is greater than 20 ft. per second in the last body, and it is therefore obvious that provisions against entrainment are indispensable. It is well to remember that there is a double danger in these high velocities, for not only is the projection high, but also the size of the drops becomes much smaller as the speed increases, making the floating effect of these greater on the principle that feathers fly in the wind more rapidly than other heavier and less bulky material.

It is generally accepted practice to provide catchalls above each body of a multiple effect even though proportionately the losses are small, as the cost of such additional equipment fully pays for itself in a short period of time inasmuch as the quantities of solution treated per unit of time are very great.

In standard effects, one can secure excellent preliminary separation before entering the catchalls by installing two semi-circular baffle plates in the vapor space, located on opposite sides one above the other. The effect of these baffles is to produce a double reversal of vapor currents at a relatively slow velocity, thus throwing the entrained drops of liquor to one side or the other, where, on coming in contact with the wetted walls of the vapor belt, they are deposited and subsequently drained below, thus relieving the load on the separator. In fact, with fairly low rates of evaporation, it is possible to secure good final separation by this means alone. It is best, however, to install the catchalls as well.

3) The other source of mechanical losses is "flashing." One very important factor in evaporator work is that in a given solution, to every pressure or vacuum, there corresponds a certain definite temperature, and the converse that to every temperature, there corresponds a pressure or vacuum. This axiom is of the greatest importance in this field of work, and cannot be emphasized too strongly.

For instance, at atmospheric pressure, water can be heated up to 212° F. If one tries to heat further, it boils and evaporates, but as long as the pressure is at atmospheric, no more and no less, the temperature will remain at 212° F. no matter how hard it is boiled. If the pressure is increased, and still more heat applied, the temperature will rise until it corresponds to the new pressure, when ebullition will again begin at this new temperature and continue without change until the pressure is again altered.

On the other hand, if water is at 212° F., and without applying heat, the pressure is raised, nothing will happen because the new pressure is higher than that which corresponds to the boiling point of the water. If

the pressure is lowered, or in the layman's terminology, if the vacuum is raised, the water, without application of heat will immediately begin to boil, more or less violently. This will continue as the depression takes place until the temperature corresponds to the vacuum or pressure imposed, this new pressure being lower than the original. If the pressure is lowered sufficiently towards an absolute of $\frac{1}{8}$ -inch of mercury, the water will actually boil until it freezes, and this is one of the methods of making ice.

This discussion is elementary but of fundamental importance. In the above example, for every degree through which the water is thus cooled, each pound gives up one heat unit and the amount of heat thus given up by lowering the pressure is represented by the total weight of the water cooled multiplied by the range of temperature change. The evaporation accomplished in this way, is called "flash."

The mechanical behavior of a mass of liquid flashing is quite different from one which is being evaporated by the application of heat through a surface. In the former case, small bubbles are formed throughout, swelling the entire mass; in the latter, bubbles are formed on the heating surface and broken off by circulation, with a relatively small amount of swelling.

The extent of this mass volume expansion depends on several factors. In the first place it is proportional to the range of temperature through which the flash takes place. It depends much more upon the specific volume of steam corresponding to the pressure under which the reaction is taking place. Thus, it is readily seen that a flash of 10° from 212° F. to 202° F. on a mass of 3000 gallons, roughly 25,000 lbs., would yield 250,000 B.t.u.'s, and a flash down from 124° F. to 114° F. would yield 250,000 B.t.u.'s.

The latent heat of steam at 202° F. is 976.6 B.t.u.'s per pound, and the specific volume is 32.33 cubic feet per pound. Similarly, the latent heat of steam at 114° F. is 1027.8, and the specific volume is 238.2 cubic feet per pound. The volume of vapor released in the first case is therefore

$\frac{250,000 \times 32.33}{976.6} = 8280$ cubic feet, or roughly twenty times the volume of the original liquid. In the other case, the figures are $\frac{250,000 \times 238.2}{1027.8} = 58,100$ cubic feet, or one hundred and forty-five times the volume of the original liquid.

These figures, startling as they are to those who have never thought about them, represent the facts, and conditions easily obtainable in a fair sized evaporator. The reaction is not instantaneous, but under adverse conditions, it may be so rapid as to cause the whole mass of liquor in the evaporator to swell up and reach the vapor outlet above, with resultant carrying over and loss, especially in the last body.

It is therefore evident that sudden changes of vacuum must be carefully avoided, especially if the amount of liquor in the evaporator at one time is considerable, as such changes are likely to be accompanied with serious losses. The only remedy when there is such a change is to break down the vacuum, and allow it to build up slowly to the desired point. Even this requires care. There will be discussed in another chapter some

of the complications to be encountered in this connection and the precautions necessary to guard against them.

At this point attention is called to the fact that in all multiple effects operating in parallel current, there is a continuous recurrence of the flashing on a small scale when the feed is passed from the first effect to the next in line which is cooler, and in designing the feed pipes, this fact must be taken into consideration, for improper attention to this problem will involve difficulties from several standpoints, as will be shown elsewhere.

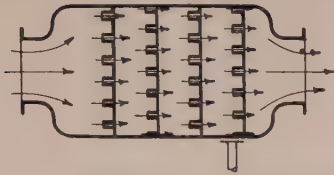
Catchalls. There are in use to-day many different catchalls whose effectiveness is more or less satisfactory, depending upon the nature of the work demanded of them.

Separator failures are due to several factors which it is well to discuss: In the first place, it is usually a fact that on account of the pressure loss in going through most separators, the vacuum in these will be sensibly greater than in the evaporator by the amount of this friction drop. Such being the case, the drain or return connection for the removal of the entrapped liquors, should be water sealed in the evaporator. If not, there is an active vapor current going up this pipe from the evaporator to the catchall which prevents free drainage, sometimes actually blowing liquor back into the separator where it is projected into the rapidly moving vapor currents, and carried away with loss. This water seal requires attention, and must be inspected at periodic intervals to make sure that it is clear at all times, otherwise the catchall becomes useless. Where possible, it might be a good idea to put a sight cross in this drain pipe so that the operation can be kept under observation.

Another fault which is quite general in some designs provided with baffles, is that such baffles are arranged in such a way that particles of liquor, after having been caught by impingement are allowed to drip in the main path of the vapor, where they are caught up again and carried away.

Below are given a few specific examples of types of catchalls or separators in common use with evaporators. Space permits only showing a few of these.

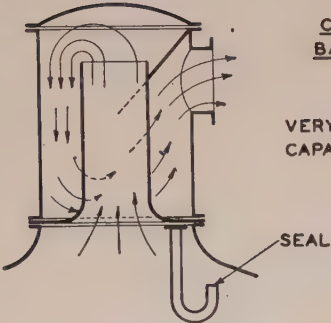
HODEK SEPARATOR.



HAVING A SERIES OF PLATES
WITH SMALL NOZZLES OFF-
SET FROM ONE ANOTHER.

EXCELLENT CATCHALL BUT WITH HIGH FRICTION LOSS.

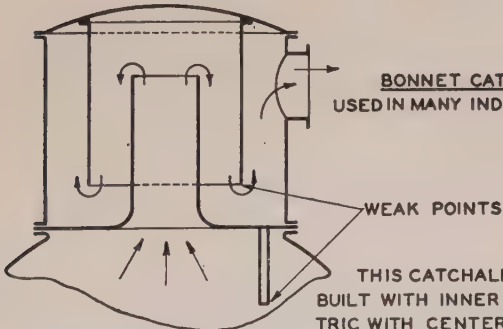
CENTER-TUBE-ANGLE
BAFFLE SEPARATOR.



VERY GOOD CATCHALL WITHIN ITS
CAPACITY LIMITS ABOUT 200FT.PER SEC,

BONNET CATCHALL.

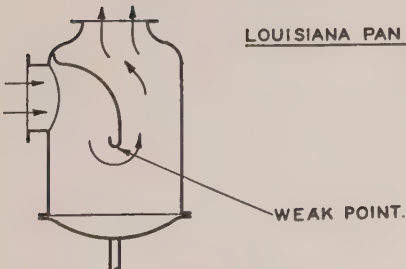
USED IN MANY INDUSTRIES.



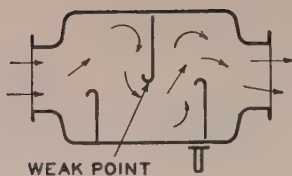
WEAK POINTS

THIS CATCHALL IS SOMETIMES
BUILT WITH INNER BONNET ECCEN-
TRIC WITH CENTER PIPE. THE
INHERENT DEFECTS HOWEVER ALWAYS EXIST TO A MORE OR
LESS MARKED EXTENT.

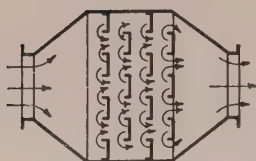
LOUISIANA PAN CATCHALL.



WEAK POINT.

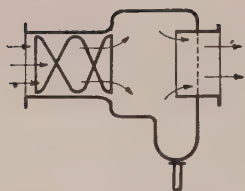


LOUISIANA EFFECT
CATCHALL.



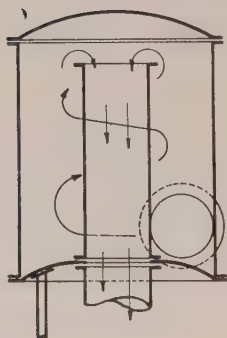
MULTIPLE CHANNEL
CATCHALL.

WEAK POINT - RE-ENTRAINED
FINE SPRAY.



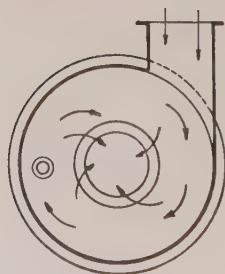
SCREW TYPE
CENTRIFUGAL CATCHALL.

WEAK POINT - CENTRIFUGAL SEPARATION
PRACTICALLY NIL IN THE
CENTER OF VAPOR STREAM.



THE WEBRE CENTRIFUGAL
SEPARATOR.

GOOD FOR ANY VAPOR SPEEDS WITHIN
THE LIMITS OF PRACTICE. THIS IS THE
AUTHORS ANSWER TO THE STATEMENT
THAT FINE PARTICLES OF LIQUOR CAN-
NOT BE SEPARATED FROM A RAPIDLY
MOVING VAPOR CURRENT. AT 250 FT.
PER SECOND, WITH LARGE QUANTITIES
OF LIQUOR GOING THROUGH, THE
ENTRAINMENT LOSSES ARE ONLY
1/50 PER CENT.



Chapter 13.

Natural Circulation.

There is perhaps no phase of evaporation that requires more careful study than circulation. Especially is this true with regard to vertical tube evaporators, which are in more general use than those of any other design. Owing to the difficulty of obtaining reliable data, the subject has never been thoroughly investigated, and most of the information in use to-day is based on pure guess work. It is well, therefore, that the matter be carefully analyzed here.

Much of the discussion will of necessity be from the theoretical point of view, supplemented where possible with actual observations. As has been emphasized in the text, such theoretical work, no matter how painstakingly carried out, will very often be seriously modified by practical considerations. It is, however, possible to use theory to determine the limits within which practice must necessarily be confined. Therefore, one object of the author in this chapter is to determine the maximum natural circulation theoretically possible in evaporators. If, as observations indicate, the usual circulation found in practice is actually far less than the theoretical maxima here calculated, that fact will only strengthen the conclusions drawn from the theory.

Before taking up the theory, it is well to obtain a clear idea of what happens in the tubes of an evaporator under working conditions. As a matter of observation, the transmission of heat in vertical tubes generally takes place in three stages, the physical conditions in each stage differing from those in the other stages.

In the first stage, which occurs at the bottom of the tubes, the liquid being evaporated is heated until its temperature reaches the boiling point. (It may be noted here that, due to the hydrostatic head, the temperature and pressure in stage I, at the point where the bubbles are first formed, will in general be greater than the temperature and pressure existing at any point above the top tube sheet.)

In the second stage, there occurs an intermediate condition. Globules of vapor are formed upon the heating surface and owing to the sluggish movement of the solution are not quickly liberated. Consequently these globules actually interpose a film of vapor between the liquid and the heating surface. In this stage, even when the bubbles are once liberated, they are quickly condensed by swirls or strata of liquor which have not yet reached the boiling temperature, for due to the flow occurring in stage I, the core of the stream will manifestly not heat as fast as the outer skin.

In the third stage the condition in stage 2 gradually merges into another, where the growth of bubbles is interfered with by the rising stream of liquid and vapor. As the speed of this stream increases toward the top of the tube, it becomes so great that the bubbles are torn off from the wall as soon as they acquire even microscopic proportions. In this stage, the temperature throughout the stream is high enough so that no vapor can be recondensed.

From the foregoing discussion it becomes evident that the coefficients of heat transmission will be very different at different points along the length of any given tube. Fig. 1 indicates the author's idea of the probable temperatures, speeds, and physical reactions occurring at various points

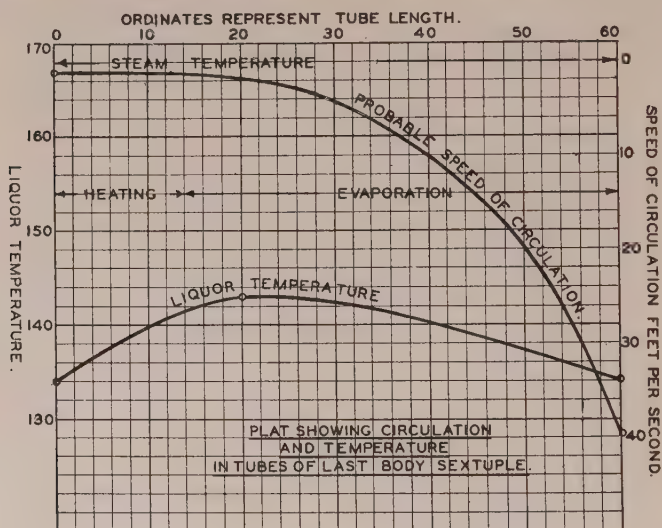


FIG. 1.

along an evaporator tube. It is based on temperature conditions known to exist in the last body of a certain multiple effect. There is no proof that this curve represents the facts, but there is good evidence supporting that conclusion. (See also Fig. 4.)

Although it is not possible with our present knowledge to establish exactly what occurs throughout the length of an evaporator tube, it is possible to establish what occurs at the two ends. Some time ago careful tests were conducted in a large plant to find out what heat transmission could be expected when evaporator tubes were used in a purely heating operation. The customary proportions of such tubes make them poorly adapted to this work, and the transmissions recorded are considerably lower than might be expected with tubes of more favorable proportions. Nevertheless, the information which is platted in Fig. 2 is valuable, as it gives a fair index of what actually takes place in evaporators in the first

stage of heating outlined above. It shows in a very striking manner the effect of varying the speed of flow of the liquid entering the tubes. The explanation of the increase of heat transmission probably lies in the breaking-up of the "surface films" by the more rapid currents. (Chapter 20.)

It is also possible to estimate the amount and volume of the vapor leaving the top of individual tubes when the proportions of these are known, together with the average rate of evaporation per square foot per hour,

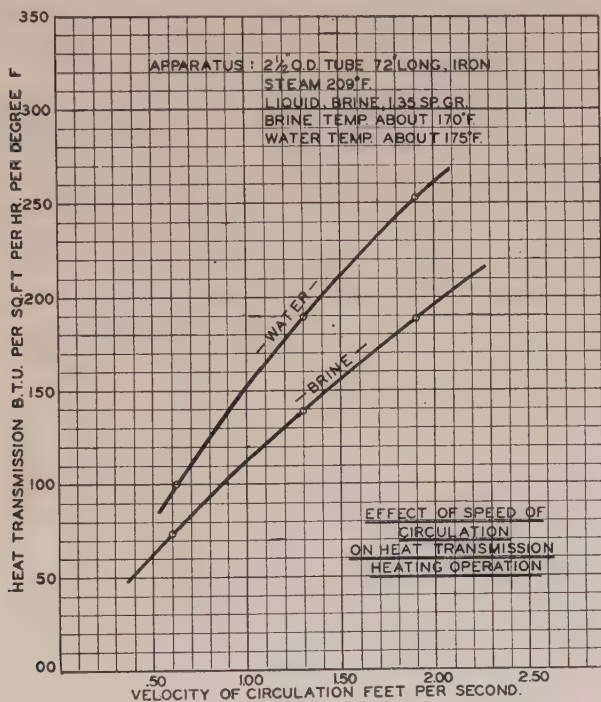


FIG. 2.

the vapor temperature and pressure, and the corresponding specific volume of the vapor in cubic feet per pound. Maximum velocity occurs when the vapor leaves the tube and evaporation is ended. Under varying conditions in practice, such velocities will range from 2 feet to 200 feet per second.

Fig. 3 gives a quick method for making the above determination. In order to use this chart, begin on the lower Y-axis at a point which corresponds to the tube diameter and proceed horizontally to the left as far as the curve corresponding to the length of the tubes. From this point proceed vertically to the curve in the upper left hand quarter of the chart corresponding to the rate of evaporation in pounds per square foot per

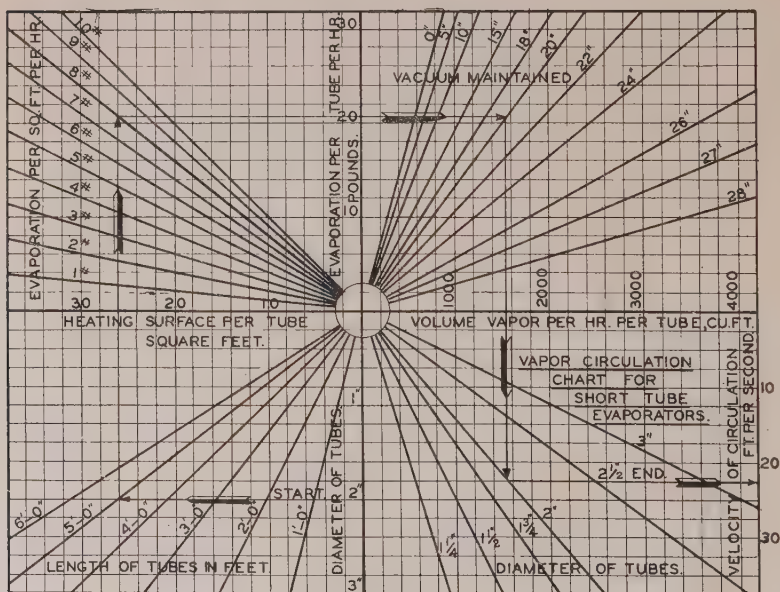
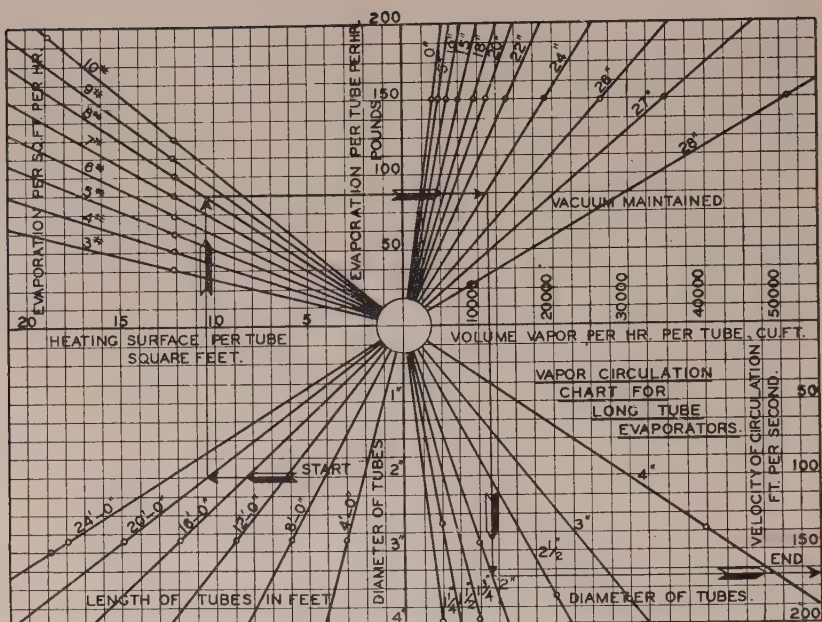


FIG. 3.

hour. Then follow to the right horizontally, intersecting the curve corresponding to the vacuum under which the evaporator is working. From here, proceed downward into the lower right hand corner of the chart, intersecting the curve corresponding to the tube diameter. From this point proceed horizontally to the right as far as the second Y-axis, at which read the vapor velocity in feet per second.

(Note: The velocities as determined from this curve will be slightly less than those actually obtained, as any variation due to the volume of included liquid has necessarily been ignored.)

The working out of a specific example is indicated by the lines and arrows of Fig. 3, as follows: A 2-inch tube, 5 feet long, has 2.6 sq. ft. of heating surface. At 8 lbs. per square foot per hour it will evaporate 20.8 lbs. of water per hour, which under 20-inch vacuum will give about 1550 cubic feet of vapor per hour, and when this vapor passes through the end of the 2-inch tubes, its speed will be 22.5 feet per second.

Up to this point only the conditions in single tubes have been considered. With this work as a basis, it now becomes possible to attack the main problems of circulation proper. The word circulation implies a complete circuit, and as applied here to evaporators it is understood as the combination of a light, vapor-swollen, rapidly rising current in some parts of the apparatus, accompanied by a corresponding slower current of liquor moving downward in other parts.

Direct measurement of circulation is a difficult task, and attempts made in the past have not given very satisfactory results. In one instance the liquor flowing back to the bottom of the evaporator was weired, but the readings were so uncertain that the only definite conclusion which could be drawn was that the velocity was relatively small. In another instance an orifice was placed in the "downtake," or return pipe, and the actual liquor speeds in stage 1 in the tubes, as calculated from the figures obtained, was found to be only about 3 inches per second. In this test the apparatus used was a long tube evaporator with tubes of 2 inches diameter, 16 feet long. The rate of evaporation was about 10 lbs. per square foot per hour and the vacuum was 26 inches. In view of the general impression with regard to velocities of circulation, the low figure obtained here was somewhat surprising.

In spite of the statements of many authors to the contrary, circulation can not be due to convection currents; for convection currents are brought about by local heating or cooling, which causes only small differences of specific gravity in the solution so heated. Such currents must of necessity be very sluggish, and convection alone could not conceivably produce velocities of 200, or even 10, feet per second.

As stated above, the real cause of circulation is the difference in specific gravity between two connected columns which are normally balanced when at rest. The heavy column consists of the liquid in the "downtake," or return pipe through which liquor returns from the top to the bottom of the evaporator. In this no evaporation takes place. The light column consists of the liquor and vapor rising through the tubes. When the evaporator is stopped, these two columns are balanced and therefore at

rest. When evaporation begins, vapor bubbles are formed under the surface of the liquid in the tubes, making this side of the circuit lighter and more bulky, so that a greater height is needed to balance the column on the other side. Then, when enough vapor is mixed with the liquor, the length of the light column becomes greater than the length of the tubes

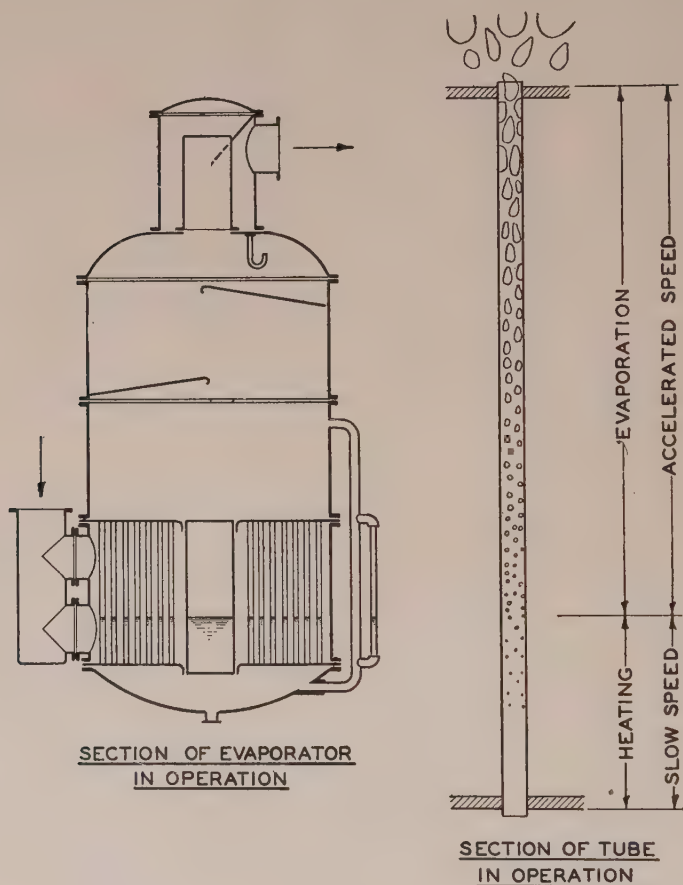


FIG. 4.

in which the vapors are being generated, and so actual circulation begins. After the mixture leaves the top of the tubes, the entrapped vapor bubbles are liberated and proceed upward in the vapor space, while the liquid returns to the bottom via the "downtake." Fig. 4 shows in cross-section an evaporator in operation, and also a large section of one tube, illustrating the physical reactions taking place. The whole operation is quite similar to that of pumping water from a deep well by means of an

air lift, in which the hydraulic gradient balances the water and air in upward motion inside the well.

As the rate of vapor formation in a tube of uniform diameter increases, the speed of the vapor must also increase. In other words, there is an acceleration. This acceleration in the tube involves a velocity head, which is balanced by a corresponding hydrostatic pressure head in the downtake. It therefore is theoretically possible to calculate the specific gravity of the liquor-vapor mixture leaving the top of the tubes under any given conditions, and consequently the amount of circulation also, provided the hydrostatic head balancing the acceleration is known. The practical difficulty in working out this calculation is that in general this hydrostatic head is not known. Fortunately, however, limiting values for the head are easily determined.

Evaporators are usually provided with water gauge glasses which show the approximate static level; that is, the approximate height of the liquid in the return column or "downtake." The static head indicated in this way is not, of course, all taken up in balancing the velocity head in the tubes. For in order to so take it up, evaporation would have to begin as soon as the liquid entered the tubes, and the preceding discussion has shown that this is not the case, besides, part of this head overcomes friction. There is considerable heating to be done before evaporation starts, and part of the static head indicated is taken up in balancing the liquid in stages 1 and 2. However, the gauge level furnishes a convenient limiting value for the static head, and any calculations based on taking the full amount as balancing the dynamic head will give a calculated amount of circulation necessarily greater than the circulation actually encountered in practice. A specific example will illustrate the method and use of this calculation.

Many standard evaporators have tubes 60 inches long, and it is customary in these to carry the levels one-third the height of the tubes. Assume last body conditions with 20-inch head as above; a vapor velocity of 50 feet per second, and 26 inches vacuum. Using the formula $V^2 = 2gH$, or $H = \frac{V^2}{2g}$, we have:

$$H = \frac{50^2}{2 \times 32.2} = \frac{2500}{64.4} = 38.8 \text{ ft.}$$

In other words, a column of mixed vapor and liquid 38.8 feet high would just balance a column of liquid 20 inches or 1.67 feet high. To simplify calculations, assume that both columns have a cross section of 1 sq. ft. The weight of the column of liquor, whose specific gravity is here assumed as 1.04, will be

$$62.3 \times 1.666 \times 1.04 = 108 \text{ lbs.}$$

The weight of a column of pure vapor, 38.8 ft. high, at 26 ins. vacuum, is:

$$0.0057 \times 38.8 = 0.22 \text{ lbs.}$$

Therefore the liquor in the mixed column weighs

$$108 - 0.22 = 107.78 \text{ lbs.}$$

corresponding to a height of 1.664 ft. with a 1 ft. sq. base.

(Note: The above calculation does not take into account the weight of vapor displaced by the liquor, as this amount, about 0.008 lbs., is so small as to be negligible.)

From the above calculation we can get the rate of liquor circulation as being:

$$\frac{\text{Liquor in mixed column cu. ft.} \times \text{Velocity of mixed column}}{\text{Total volume of mixed column at top of tubes.}}$$

$$\text{or } \frac{1.664}{38.8} \times 50 = 2.15 \text{ ft. per second.}$$

Reference to Fig. 2 shows that the probable coefficient of heat transmission for the above speed is about 200. This is for clean tubes. As the tubes foul rapidly at the bottom, the actual coefficient during the heating operation would be on an average half that value, or 100.

Similar calculations for varying vapor speeds show positively that as the vapor speeds increase the liquor speeds decrease to correspond. For example, with a vapor speed of 150 feet per second, the liquor speed can not be over 0.7 foot per second with a 20-inch head. This gives a coefficient of *H.T.* equal to 80 for clean tubes, and 40 for foul tubes. The above figures of course show only the maximum rates of liquor circulation theoretically possible, the actual liquor speeds obtained in practise are much smaller.

A series of tests made by the author some years ago are of interest as indicating the amount of static head used up in balancing the liquor in stages 1 and 2. In these runs the height of the levels as indicated in the level gauges was varied, while other conditions remained the same. It was found that the apparent static head reached nearly to one-third the height of the tube before any circulation at all was produced.

With a level one-quarter the height of the tubes, it was found that the upper ends of the tubes remained dry, and the thermometer indicated a slight superheat in the vapor. Under these conditions, of course, no liquor was coming back through the downtake, and the flow into the tubes was simply a replacement of vaporized liquid. The maximum heat transmission was secured when the ends of the tubes were kept wet, a condition obtaining when the level was slightly over one-third the tube height. When the level was raised still more, the circulation became observable, and the rate of heat transmission dropped slightly.

The results of the above experiments can not, of course, be applied directly to other evaporators, working under different conditions. It can, however, be stated definitely that for every vertical tube evaporator, working under any given conditions, there is a certain critical gauge level, below which no circulation takes place. This level, when determined, will give

a good measure of the static head which balances stages 1 and 2 in the tubes. It is the experience of the author that in usual practise the "critical level" is at least one half of that indicated by the gauge glass. When cold liquors are continuously fed into the evaporator from below the tube sheet this level will be considerably higher, as more of the tube length is then taken up in heating.

It is therefore safe to take the average circulation found in practise as not more than one half of that calculated above as the maximum possible. Thus in the examples given the actual circulation for 50 feet vapor speed would be about 1 foot per second, and for 150 feet vapor speed only about 4 inches per second. The actual coefficients of $H.T.$ of course are also correspondingly less.

The following conclusions may be drawn from the foregoing discussion:

First: It is unwise to do any considerable amount of heating in an evaporator, and doubly unwise in long tube evaporators, where the vapor speeds are high. (For this reason, Kestner, in his original equipment, provided preheaters so as to start up the tube with a flash instead of a heating operation.) If conditions are such that heating is unavoidable, the feed should be introduced into the vapor space by means of sprays, so that heating can take place by direct contact with the vapors. The apparatus can take care of the additional evaporation load efficiently, as this is what it was designed for.

Second: In evaporators with high vapor velocities, especially long-tube evaporators, there is a tendency for the bottom of the tubes to foul rapidly, owing to the sluggish flow of the liquid. Under conditions of this sort, it is wise to keep the gauge levels high enough to insure a fair circulation, even though lower levels give a temporarily better evaporation.

Third: The usual downtakes in standard evaporators are much too large, and could be reduced very materially without in any way interfering with their efficiency. This is doubly true in the case of long-tube evaporators.

The reader is again referred to plat Fig. 1, which shows our general conclusions as to the physical reactions that occur in a single tube. Fig. 4 should be studied at the same time, and in this way it will be possible to form a good picture of the problems involved.

The previous discussion has been confined to the velocities of circulation of liquor and vapor in evaporators. There are, however, other points to be considered.

Some time ago, the author conducted experiments for the purpose of observing the physical reactions occurring during evaporation, and their influence on the transmission of heat. A small evaporator such as is shown in Fig. 4 was filled with cold water flush with the upper tube sheet. Steam was then admitted from a constant high pressure main until the pressure in the steam belt registered 5 lbs. gauge, and the air was removed from the calandria. The reactions proceeded without the setting of the valves being changed, and were recorded as follows:

The vapor space being open to the air, as the temperature of the water

gradually increased and the pressure in the steam belt increased also, finally reaching 10 lbs. gauge just before ebullition started. Toward the end of this period, a large quantity of bubbles could be seen attached to the inside of the tubes. These bubbles probably consisted of a mixture of air and vapor, and undoubtedly acted as an insulant between the water and the heating surface, thus slowing up the transmission of heat. There was a slight circulation due to convection, the water in the tubes being hotter than that in the downtake. The succeeding changes came about very quickly as follows:

First, ebullition began.

Second, the pressure in the steam belt dropped back from 10 lbs. to 5 lbs., though the rate of admission of steam remained constant.

Third, the level as indicated by the water gauge glass dropped to a point about half way up the tubes, while the level of the frothy mixture of vapor and liquid rose to a point about 12 inches to 15 inches above the top tube sheet. It was noticed that the discharge of the tubes nearest the downtake interfered very materially with the horizontal flow of water from the outer periphery towards the centre of the evaporator, thus impeding circulation.

After these observations, steam was turned off, and the levels as shown in the gauge glass came back to the original point.

Consideration of these phenomena yields interesting results. As the barometer was at 30 inches, the vapor temperature was at 212° . The steam temperature at 10 lbs. pressure was 240° F., and at 5 lbs., 227° F. Consequently, after ebullition started, the temperature drop through the heating surface was 15° , while before ebullition, it was from two to eight times as great, according to the stage of reaction. During the heating operation, however, the circulation was entirely due to convection and therefore was very sluggish, whereas afterwards, it was due to evaporation, and, therefore, much more rapid. The author has shown previously the result of such a change, and the importance of maintaining an adequate circulation. (See Fig. 2.) Towards the last, the insulating effect of the bubbles was also undoubtedly very large.

The change in gauge level may be explained by the fact that some water was apparently held in suspension by the impingement of the discharge from the tubes. Also a considerable part of the liquid was resting on the flat portions of the tube sheet, and consequently was not recorded in the water gauge glass. The lesson to be learned from the difficulty of flow is that the horizontal distance to be travelled by the liquid above the tube sheet must not be too great. Consequently, it is better, especially in large evaporators, to have a number of smaller downtakes distributed in the steam belt, instead of a single large central opening.

Chapter 14.

Mechanical Circulation.

In the previous chapter, the various aspects of natural circulation in evaporators have been discussed. There are a number of problems, especially in the chemical industries, where it is impossible to secure satisfactory results without mechanical, or artificial circulation.

Such is the case when a saturated solution must be evaporated for the purpose of crystallizing out the salts in solution. Here, as soon as evaporation takes place, a corresponding amount of crystals are precipitated out of the solution. This would not affect the apparatus adversely were it not for the fact that with natural circulation, these crystals are formed where the evaporation takes place, namely, on the heating surface, and in spite of all attempts to prevent it, a large quantity of these adhere to the tubes, soon forming a layer that grows thicker and thicker, making the transmission of heat more and more difficult, until within a very short time, depending upon conditions, further operation is possible only at such reduced rates, that it becomes necessary to shut down for a washout.

So far as heat transmission is concerned, these salts accumulating upon the surfaces of an evaporator act in about the same way as any other accumulation of scale, and the reader is referred to the chapter on scaling and fouling for further elaboration.

Such coating of surfaces will be more serious as the exposed side of the tubes is rougher, and therefore offers more crevices and attachment spots for such accumulations. It is therefore advisable to have tubes whose surfaces are as smooth as possible to avoid the difficulty.

In addition, sometimes, the crystals are very hard and sharp, and cause serious wear in the tubes, particularly near the top ends, where the speed is highest, the effect being very similar to a sand blast.

Most salts encountered are more soluble hot than cold, and if the solution is merely heated, there will be no precipitation; and it is therefore possible to secure fair heat transmission in such a purely heating operation. Fig. 1, which is an elaboration of Fig. 2 in Chapter 13, shows heat transmissions obtainable in this way, and indicates that it is easily possible to do better than 200 B.t.u.'s per square foot per hour per degree drop with saturated brines, where the speed of circulation is 2.5 feet per second, which is well within the limits of practice.

If the transmissions recorded in the operation of the evaporator with natural circulation are less than this figure, then by using the surface for a purely heating operation with mechanical circulation, better results will

be obtained. It is easily possible to make an evaporator perform its work of heat transmission in a purely heating operation by the intelligent use of mechanical circulation. This is usually accomplished by installing a propeller in the downtake, which acts as a pump, forcing the solution down the centre and up through the tubes. The amount of head required to create this flow is really very small. Fig. 1 shows it for $2\frac{1}{2}$ inch tubes 6 feet long. These data were secured by actually flowing brine through a given tube of the dimensions stated, and observing the total pressure drop between the two ends for various velocities. At the same time heat

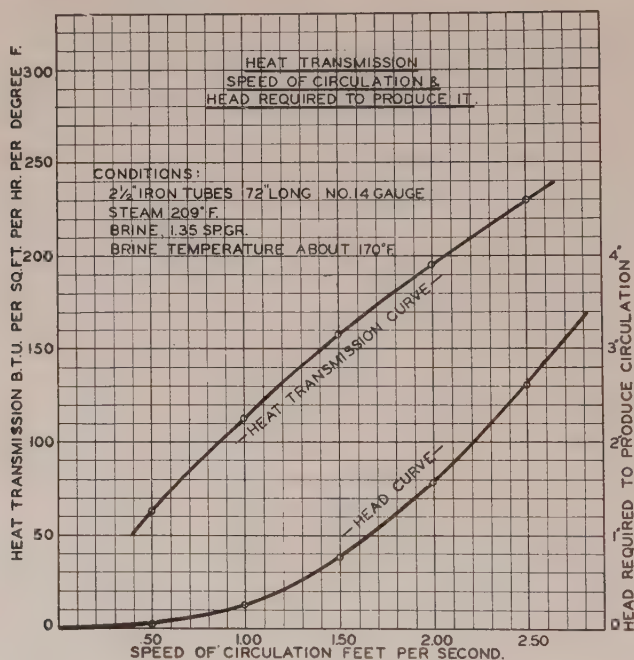


FIG. 1.

transmission observations were made, and it is felt that this represents the facts accurately, as all observations were checked and rechecked many times.

When an evaporator is operated with mechanical circulation, the level of the boiling liquid inside must be maintained *well above the upper tube sheet*, probably 18 to 24 inches. Fig. 2 shows a cross section of an evaporator equipped with mechanical circulation, and represents a type in general use in the salt industry. Such evaporators are built in very large sizes, even up to 30 feet in diameter.

In operation, the reactions follow a definite cycle, as follows: The saturated brine starting up from the bottom is heated as it goes through

the tubes, and, the solubility of the salts in solution being greater hot than cold, there is no precipitation. The increase of temperature of the material passing through the tube is very small. For instance, with a 2 inch

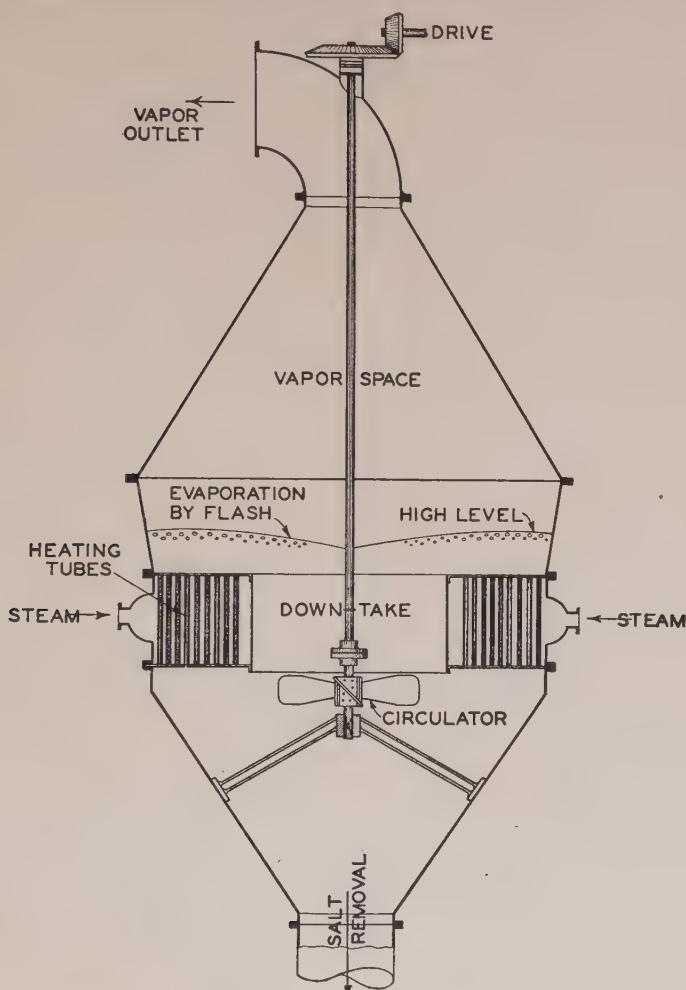


FIG. 2.—Section of Evaporator with Mechanical Circulation.

tube 6 feet long, working with a net temperature drop of 20° F. between steam on one side of the surface, and brine on the other, the circulation being 2.5 feet per second, and the transmission 200, the heat transmitted per hour per tube would be $3.14 \times 200 \times 20 = 12,560$ B.t.u.'s and per second 3.5 B.t.u.'s. The inside cross section of the tube is 2.76

sq. in. (the outside diameter being 2 inches, and the inside $1\frac{7}{8}$ inches). The volume of brine passing through per second is therefore 82.5 cu. in. Assuming the specific gravity of the brine at 1.3, the weight per second is $\frac{82.5}{1728} \times 62.3 \times 1.3 = 3.87$ lbs. The specific heat of the brine is about 0.85, and the rise in the temperature will be $\frac{3.5}{3.87 \times .85} = 1.065^\circ \text{ F.}$

The hydrostatic pressure above the upper tube sheet will make evaporation in the tubes impossible for such a small rise, and it will be only after the brine reaches the surface of the liquid that evaporation will take place in the form of flash. Actually, the brine could be appreciably supersaturated in the tubes without precipitating, as there is a definite lag to this reaction. No doubt, some of the brine is shunted to the downtake and carried to the bottom without ever reaching the boiling surface. This is diffused with the rest of the solution coming up for the next circuit.

The observed temperatures, as compared with the theoretical, show that the brine is practically always superheated above the top tube sheet, and sometimes much more than one would expect.

As evaporation occurs, so will the salts precipitate, and the mixture will now be pumped into the bottom by the circulators, where, due to the semi-quiet conditions existing, the heavier crystals settle to the bottom whence they are removed. The brine with light crystals makes a new circuit through the heating surface.

This procedure looks very simple, and, from the theoretical standpoint, nothing could be more so, a difficult evaporator problem having been converted into a simple heating problem. In practice, there are many pitfalls which must be guarded against.

Presumably, there is no reason why salt should accumulate on the heating surfaces once mechanical circulation is properly installed, and one would certainly be justified in expecting a heat transmission of 200. Salt does accumulate, and the expected transmission is not realized. The owner, on the other hand, does not accept excuses, so that the practical end of the proposition has to be carefully investigated.

There is always a good reason for everything, and it is felt that this deviation from the theoretical is due mainly to defective design of apparatus which does not accomplish what it was intended to do. There is no doubt that the average mechanical circulator does not provide sufficient circulation, which is very difficult to measure, and therefore elusive. Furthermore, even if such a determination is made, there is no proof that the brine is equally distributed among all the tubes.

In one instance the actual velocity of the liquid was measured as it flowed through the tubes of a large evaporator so equipped, and instead of having a speed of 2.25 feet per second, for which the design was made, it was only 0.66 feet per second, showing about 75 per cent. slip in the propeller. This latter velocity was measured at the outermost tubes, and when it is remembered that there is an inevitable swirl imparted to the brine in the lower part of the apparatus, creating a centrifugal action, it is no doubt true that the velocity recorded was much greater than the

average, and that there is a fair possibility of there being no circulation at all in the tubes near the downtake, or even a negative circulation, as indicated by the irregular boiling above the top tube sheet.

It has also been established that suspended salts in large quantities materially interfere with the transmission of heat, and that there is much to be gained by their complete removal as soon as formed.

It is therefore evident that much care and study must be expended on the proper design of all such apparatus if proportionate returns are expected. The pumping operation in this work is radically different from other pumping operations, involving very large volumes of liquid per unit of time, but very small heads, as has been brought out in Fig. 1 referred to before. Such being the case, in designing the circulators, in order to secure a fair measure of efficiency, the velocity heads must be recovered by the use of venturi tubes, hydrocones, or some such arrangement, as this velocity head is probably much greater than the actual head required to force the liquid through the tubes. The swirls in the discharge of the pump must be rectified so that the brine as it approaches the tubes from the bottom moves in an easy smooth-flowing vertical direction.

If, in addition, the surface tension of the liquid being evaporated is low, there will be a decided tendency to foam, which would not be noticed with natural circulation. If an attempt is made to overcome this trouble by lowering the levels, and operating with natural circulation, the tubes will quickly salt up.

As to the velocity of circulation advisable in such evaporators, the range is quite limited, for the power expense increases enormously when certain limits are passed, as can be readily seen in Fig. 3, which represents an estimate of the power requirements for circulation in a very large evaporator. The selected working speed in this case was 3 feet per second, under which conditions a transmission of 260 net is expected with brine.

The experiments reported in Fig. 1 were made with $2\frac{1}{2}$ inch tubes, as stated before, and these proportions are certainly not the best for a heating operation. It would probably be much better to use smaller tubes, so as to reduce the core effect, and skin heating which was apparent and recorded at the time. For this reason, designers of surface condensers use small tubes, even $\frac{5}{8}$ inch, for liquor heating and steam condensation are one and the same thing.

In addition, a reduction of tube diameters permits a reduction of the size of the evaporator for a given surface, thereby decreasing the first cost.

The determining factor as to the use of mechanical circulation as against natural circulation in evaporators will be the overall coefficient of heat transmission obtained by one method as compared with the other, based on the whole period of operation from washout to washout. The additional first cost, cost of operation, and complication must be charged up to mechanical circulation, and unless there is a decided advantage, it is better to use natural circulation.

It is possible and practical to evaporate crystallizing solutions by the

use of natural circulation, and there are many evaporators so operated for this work. A number of precautions are necessary. First, it is advisable to use large diameter short tubes, and to carry high levels in order to have good liquor circulation. The high levels will hold down the heat transmission at the beginning of the cycle, but in the long run more average work per hour will be accomplished than if the levels had been maintained

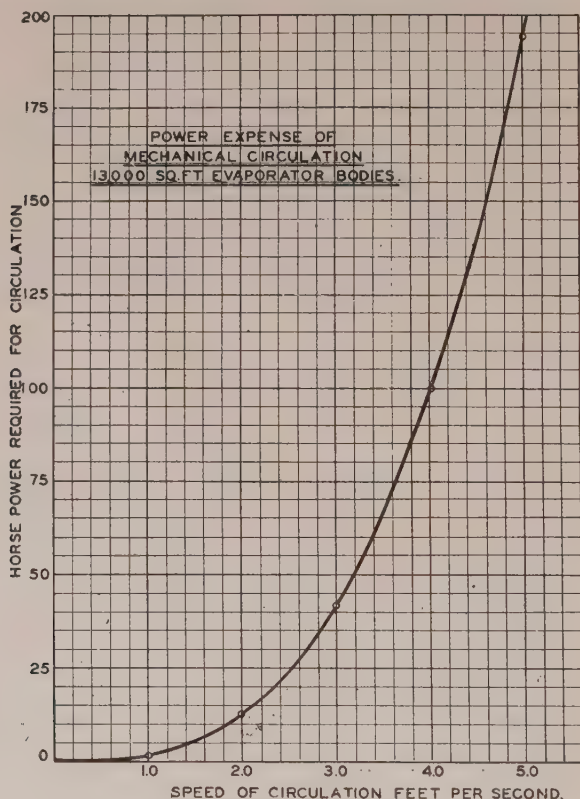


FIG. 3.

low. In the latter case, there will be a rapid accumulation of salt on the surfaces with consequent depreciation of heat transfer, and the necessity of a shut down for cleaning or operation at a very much reduced rate.

It is absolutely impossible to evaporate certain solutions without mechanical circulation, as has been demonstrated on many occasions, and the characteristics of the material must be taken into consideration in arriving at the proper decision.

In designing for this work, it will be well to follow surface condenser practice, which has been thoroughly investigated, and is therefore well

established. The big difference will come from the fact that the rise in temperature of the solution upon going through the tubes will be much less, and therefore involves the use of one pass instead of several.

It might be advisable to separate the two operations completely, having heating elements with their circulation discharging into "flash chambers" of sufficient size and proper design, so that the salt crystals when formed will be permitted to settle without interference from swirls and eddy currents such as are inevitable in the ordinary design. This scheme involves a number of different problems, however, and one must therefore study it very carefully before going too far.

Chapter 15.

Selection of Tube Proportions.

In the introductory chapter, the statement was made that most of the data used in the design of evaporators were based on rule of thumb. This is especially true with regard to the selection of tube diameters and lengths.

There are advocates of all sorts of proportions, some preferring large diameter short tubes; some, long tubes of relatively small diameters, and still others advising intermediate proportions. It is true that the information available has not been thoroughly analyzed and segregated, and this probably accounts for wide divergence of opinion on the subject. Furthermore, precedent has a great influence in such matters, and will, as a rule, be the determining factor so far as local consideration is concerned. However, there should be well matured, definite reasons for such selections. Insofar as the available data permit, an attempt will be made to establish a basis for at least the determination of tube diameters and lengths in evaporator work.

The problem will be discussed from different angles. Perhaps the first thing to be considered is the matter of mechanical strength. When proportions reach extremes as to length, there is a well-defined limit beyond which it does not pay to go, for the tubes are difficult to ship and handle, are easily bent and injured, and are sometimes hard to get. Some long tube designs have lengths as great as 27 feet. Such extremes should be discouraged, and, so far as the direct influence on the evaporator is concerned, the same results can be obtained by reducing the tube diameter and the length as well. Thus it is easily possible to keep the length under 20 feet, a convenient size for shipping and handling.

So far as the strength of the tubes and their ability to withstand service are concerned, it depends upon the design of the evaporator and the way in which they are installed. There is no doubt that the short length, large diameter tube is the strongest, but this factor need be given only small consideration, as practically any size tube would be strong enough. This is especially true, if the evaporator is of the vertical type. If it is a horizontal apparatus, and the tubes are very small in diameter in proportion to their length, it will be necessary to support them in between the tube heads in one or more places. Lack of attention to this detail will result in a sag which will tend to loosen the ends, and, in addition, will hold back water of condensation, thus blanketing off heating surface.

Another bad feature of unsupported horizontal tubes is that when the

evaporator is in operation, they will vibrate seriously, resulting in leakages at their ends. So that, with proper supports in horizontal tube evaporators, the factor of tube strength should not be the one to determine the tube proportions.

If the tubes exceed a certain length, in either horizontal or vertical designs, it is necessary to provide for expansion and contraction by either installing an expansion joint in the shell, or putting stuffing boxes at the tube ends.

The greatest temperature strains in evaporator tubes occur when a hot apparatus is shut down, liquidated,* and the liquor side filled with cold water. This may be necessary in washing out for an inspection, or in an emergency for making repairs. What really happens is that the thin tubes are cooled off very quickly and suddenly, while the shell of the steam belt, being lagged, cools off very slowly by radiation, and will still be very hot when the tubes are already cold, thus creating a very serious differential contraction.

This reaction puts all the tubes under tension, and is likely to loosen them when normal temperatures are resumed, unless provisions have been made against this contingency as suggested above.

With tubes expanded on both ends, and with neither expansion joints nor stuffing boxes, a condition encountered in all standard effects, particularly if the tubes are rather long, sudden temperature strains such as described above must be carefully avoided, for, aside from theory, practice has clearly shown that such treatment will loosen the tubes.

As to the effect of tube proportions upon heat transmission on the steam side of the apparatus, it seems to us that there can be little dispute, as the determining factors are more the general arrangement of the heating surface, the grouping of the tubes, and the method of admitting steam to the evaporator rather than the individual size of the tubes. This matter is the subject of a separate chapter.

For vertical tube evaporators, the operating conditions on the liquid side are the chief determining factor. As far as heat transmission is concerned, the information available points towards the small diameter long tube rather than the large diameter short one; for, inasmuch as the actual surface of the tube is proportional to its diameter and length, and the cross section of the same tube is proportional to the square of its diameter, a given rate of evaporation per square foot of surface will give a velocity of circulation that varies inversely with the diameter for any given length. In other words, the longer the tube and the smaller its diameter, the greater the speed of vapor and liquor passing through the top end, and the faster the average velocity. Heat transmission, being a function of this velocity, will therefore be better with smaller diameter long tubes than with large diameter short ones.

It has been claimed that when the length of the tubes exceed certain proportions, it is impossible to prevent entrainment, this being doubly true, if the tubes are of small diameter, for with these fast circulations, the drops of liquid are so small that they cannot be separated from the

* A technical term in general use for emptying all the liquid from an apparatus.

vapor currents leaving the apparatus which carry them away with eventual loss. It has been positively proven that with good separators there is no danger of loss, even with vapor velocities at the tube ends as high as 200 feet per second, so that this feature can be left out of account.

The critical point of construction, where high circulating speeds are used is in the separator or catchall used with the apparatus. Its function is to catch all particles of liquor mechanically entrained in the vapor as it leaves the body of the evaporator. If this separator is not equal to the task, naturally serious losses will result, and this is not at all uncommon. The separator to be used for this problem must be very carefully designed and ample in proportions. The centrifugal type is very well adapted, and has been used successfully for this work. (See chapter on Entrainment and Its Prevention.)

It is felt, therefore, that the matter of entrainment need not be the limiting consideration in the selection of tube sizes.

The real determining factors are first, the clogging tendency of the tubes, which will be greater with small tubes than with large ones, for a given thickness of dirt accumulation will encroach more upon the cross section of a small tube than a larger one. If the nature of the solution being evaporated is such as to form deposits of scale on the surface, then the use of small diameter long tubes will invariably give trouble, and should not be permitted. This consideration is usually the most important one, for once tubes become clogged so that water will not pass through, it is a very difficult matter to clean them out, and sometimes impossible, necessitating removal and replacement by new tubes.

The second factor of importance in this consideration is one of circulation, not simply vapor circulation, but simultaneous liquor circulation as well, which is almost as important. In the chapter on natural circulation, great pains were taken to show that a definite relation exists between the rate of circulation or discharge from the top of the tubes, as compared with the rate of inlet or entrance velocity of the liquor at the bottom. Roughly speaking, for any given level of liquor in the tubes, theoretically, the product of the vapor outlet velocity by the liquor inlet velocity tends to be constant. In other words, high outlet velocities are necessarily accompanied by low inlet velocities, and conversely. In view of the fact that there is invariably a necessary heating effect, as the liquor enters the tubes from the bottom, with low liquor velocities, this heating will be inefficiently done. Also, with low speed of circulation here, scale and dirt will tend to deposit much more rapidly than otherwise. This is shown by the fact that tubes of any proportion always foul more rapidly at the bottom than at the top, and this is actually much more pronounced as the diameter of the tube is decreased, and its length increased.

There is a limit to sluggishness of liquor speed at the bottom of the tubes which it does not pay to overlook. Naturally, the nature of the solution being concentrated will determine what this limit is.

If the solution is of a crystallizing nature, it is very advisable to provide fast liquor circulation and slow vapor circulation for two reasons. In the first place, with slow liquor circulation, and the accompanying fast

vapor circulation, the crystals formed will tend to be very fine, and difficult to settle. In addition, if the vapor speed exceeds a certain limit at the top, there will be considerable abrasive action inside the tubes at the top. At the bottom or midway, there will be a deposition of salts inside, with depreciation of heat transmission. With crystallizing solutions, it is customary to carry high levels, in order to increase the liquor circulation, and this must be assisted by limiting the vapor speeds, otherwise, one will vitiate the other.

These vapor speeds are determined by tube proportions and operating conditions. Thus, by proper selection, any combination can be secured, to suit the requirements of the case in hand.

With viscous materials, it is impractical to use high speeds, and the tubes have to be proportioned accordingly.

For foamy, non-crystallizing materials, high vapor speeds are desirable, for it has been found that under ordinary conditions, no foam will be formed if the speed at the tube ends exceeds a certain critical speed of from 10 to 14 feet per second, whereas it may be very serious at very sluggish speeds of 4 to 6 feet per second.

For delicate materials, high liquor speeds are also advisable as the ill effects of temperature are reduced to a minimum, as the time element is cut down tremendously, the contents of liquor being less for long than for short tubes.

There is another decided advantage from another standpoint in favor of small diameter long tubes and fast circulation, in that it permits the installation of a much larger amount of more efficient heating surface in a given body, thereby reducing the first cost per square foot, and also minimizing the heat losses from radiation inasmuch as these are proportional to the surfaces exposed.

As to the practical advisable limits for tubes, there is no other way than actual trial to fix them absolutely, and such limits will vary according to conditions in different places and different solutions. As far as prevailing practice is concerned, tubes of many lengths have been used, from 3 feet to 27 feet, and diameters from $\frac{5}{8}$ inch up to 5 inches. For vertical evaporators, tubes smaller than $1\frac{1}{4}$ inches will probably give trouble, not only from plugging with dirt, but in expanding and handling on account of their fragility.

The large tubes of short lengths are used for sugar vacuum pans operating with viscous crystal laden solutions. Long tubes should be used for foaming liquors. Much to the layman's surprise, the first cost per square foot of evaporators with 16-foot tubes is more than those having tubes 5 or 6 feet long. In horizontal tube evaporators, the sizes vary from $\frac{3}{4}$ inch to $1\frac{1}{4}$ inches, the latter being a favorite on account of its greater rigidity. Horizontal evaporators of the Lillie type are best built with much larger tubes, probably 4-inch.

Chapter 16.

Types of Steam Belts.

In other chapters, it has been pointed out that the steam side of the heating surface of evaporators involves considerations parallel to those encountered in surface condensers, the main difference being, as a rule, that the temperature existing during the process of condensation is very much higher, and the steam density greater correspondingly.

The serious problem, and the one which has been the most obstinate to overcome, is the removal of non-condensable gas. (See Chapter 7, Sec. II.) In the first place, steam or vapor, as it enters the calandria or heating chamber, is never pure steam, but is always accompanied with a small percentage of non-condensable gas which may come from any one or all of three sources:

1. It may consist of air which was originally dissolved in the water which was fed to the boilers for steam generation. The percentage by volume of this air in steam is necessarily very small, for the solubility of air in water is low, amounting to perhaps from 1 to 4 per cent. by volume, according to its temperature and saturation. If this boiler feed has been heated in an open heater, before introduction into the boiler, most of the air will have been driven out. If boiling condensate from the evaporator is fed into the boilers, it will contain practically no air, so that the figures represent the worst conditions.

Even assuming that the boiler feed does contain as much as $2\frac{1}{2}$ per cent. air in solution by volume, and that its initial temperature were 60° F., when this is converted into steam at 5 lbs. gauge for use in the evaporator, 1 cu. ft. of water will produce approximately 62.3 (weight of 1 cu. ft. water) $\times$ 20.44 (sp. vol. steam at 227° F.) which gives 1273 cu. ft. and the air will be dissolved therein. As to the removal of this air, its volume will be greatly increased due to vapor saturation, and rise in temperature. Assuming first body conditions in which the temperature of the boiling liquid inside the tubes will be about 212° F., it will be practically impossible to remove the air with accompanying vapor at a temperature much less than 5° higher than this figure, or 217° F.

The original volume of air, 0.025 cu. ft. at 60° F. must be corrected for all these changes. The absolute total pressure in the steam belt is 5 lbs. plus 14.7 lbs., or 19.7 lbs. abs. The vapor pressure corresponding to 217° F. mentioned above is 16.22, and the residual air pressure is the difference between the two, or $19.7 - 16.22 = 3.48$. The corrected volume of orig-

inal air is $0.025 \times \frac{14.7}{3.48} \times \frac{460 + 217 = 677}{460 + 60 = 520} = 0.138$ cubic feet, or about 0.011 per cent. of the original volume of steam to be condensed.

If the pressure in the steam belt is reduced, the volume of the mixture is increased in proportion, making the residual volume of vapor saturated air to be removed greater, and its bad effects more serious.

2. As a rule, air coming from the evaporation of water into steam in the boiler plant is not the main source of non-condensable gas, for a much larger volume enters the system via the calandrias. As these are operating at a pressure below that of the atmosphere, air leakages in the various joints, valves, stuffing boxes, etc., are unavoidable. The volume of these leaks depends upon the particular apparatus under consideration, the operating conditions, and the care with which it has been installed.

3. Sometimes, there are non-condensable gases dissolved in the solution to be evaporated, and these are liberated by ebullition, and find their way mixed with vapor into the calandria of the succeeding body.

Whatever the source of these gases, they are always present, and their influence upon heat transmission must be carefully studied. It was mentioned in Chapter 5 that very often they are of a corrosive nature, and therefore attack the material in the tubes, and from this standpoint alone, their thorough removal at the earliest opportunity is a matter of prime importance.

The effect of the non-condensable gases upon heat transmission is very marked; for, assuming that a given tube is surrounded by a mixture of steam and gas in a quiescent condition such as obtains in many evaporators, the steam part of the mixture condenses upon the heating surface, leaving a residual film which acts as an insulator through which heat must pass in order to reach the tube. Thus the transmission is materially retarded, and for each additional quantity of steam condensed, there corresponds an additional residual quantity of gas which eventually forms a mixture with steam whose temperature approaches more and more that of the liquid being evaporated on the other side of the heating surface, thus ultimately stopping the passage of heat. These facts have been well established by various investigators (Kerr, *La. Bul.* 149), and must strictly be taken into consideration.

In some of the earliest literature on evaporation (Von Wykoff) it was found that the designers were hopelessly blocked in their work by their inability to solve this problem. Their practice was periodically to lower the vacuum until all the calandrias were under pressure, thus enabling them to force the air out to the atmosphere through outlets provided for this purpose, known at that time as purge cocks, and still provided in evaporators as built about twenty years ago. After proper purging, these cocks were closed off, and the operation resumed for another run, until the accumulation of gas had become such that the operating rate was so reduced as to necessitate another venting.

The heat transmission under these conditions was necessarily poor, and the temperature differences required for operation large, thus limiting the number of bodies to their multiple effects, and making impossible high

economies as practiced to-day. These considerations are basic, and apply to all designs, and certain provisions are indispensable for the proper solution of the problem.

It is advisable so to arrange the heating surfaces as to compel steam or vapor to follow a predetermined path through the steam compartment, and so to proportion this path as to obtain a high velocity of travel, the higher the better, for the purpose of scrubbing off the residual gas film, thus exposing surfaces to the full benefit of steam. The residual non-condensables are thus driven forward and replaced with rich steam, being finally removed from the end of the path by means of an air pipe which is kept open for the purpose of preventing any accumulation. One should not be afraid to vent amply, for the maximum possible heat losses from this source are really very small, as can be readily estimated from a study of the flow of steam through small orifices. Even if there is a small waste, this expense is fully repaid by better operation and minimized corrosion.

As to the limits of speed of vapor travel through the surfaces, friction losses are the determining factor, and even if moderate losses are involved, they also are fully justified by higher heat transmission. It seems that preferably, the flow of steam should be at right angles to the tubes in order to secure a better scrubbing effect. It is true, however, that very good results have been obtained otherwise. In horizontal effects, the travel of steam must of necessity be parallel to the tubes, inasmuch as it flows on the inside.

The velocity should be high throughout, if possible being faster at the end of the path than at the beginning, in view of the fact that the proportion of gases is greater at this point. In practice, in vertical tube effects, speeds have been used up to 50 ft. per second with good results, and the losses of pressure due to friction in well-designed steam belts even with these high speeds is really very small, not exceeding $\frac{1}{4}$ in. of mercury, which may be considered negligible. With such velocities, there exist practically storm conditions on the steam side, and the scouring of the gases is very thorough.

With horizontal tube effects, it is advisable to have several passes horizontally, with a good pitch to the tubes in the direction of the flow, in order to insure proper drainage of condensate. If only one pass is used, the velocity will be practically zero at the outlet end of the tubes, although it may be very fast at the inlet, and, in fact, the problem here is to keep down the velocity on account of the relatively small area of the inside of the tubes as compared with their total heating surface. For a given length of travel, however, with a given speed, the friction is undoubtedly less than for vertical tube effects on account of the uniformity of direction, and smoothness of walls.

As a specific example of horizontal tube designs, consider the last body of a quadruple effect having 480 $1\frac{1}{4}$ -in. tubes 14 ft. 0 in. long, giving a total of 2200 sq. ft. of heating surface. Presuming that this apparatus is condensing six pounds of steam per square foot per hour, and that the vacuum in the steam belt is 18 in. referred to a 30-in. barometer, the total

steam condensed per hour will be 13,200 lbs., and per second, 3.67. The specific volume of steam at this vacuum is 63.3 cu. ft. per pound, and the volume of this steam would therefore be 232 cu. ft. per second.

The inside diameter of the tubes, assuming them to be No. 18 B.W.G., will be $1\frac{1}{8}$ in., and the area corresponding will be about one square inch. If there is only one pass, the speed at the entrance will be in feet per second $\frac{232 \times 144}{480} = 69$, which will dwindle down to zero at the outlet end of the tubes, the average being 34.5.

Assuming that the bank of tubes is twelve high, and forty wide, a number of multipass arrangements can be made, and below is given a table of the most likely combinations. In the three pass design, the minimum speeds are over 10 feet per second, except in the last pass, which represents only a small proportion of the total heating surface. This scheme should therefore yield excellent results.

METHOD OF GROUPING TUBES FOR HORIZONTAL EVAPORATORS.

FOR SINGLE AND MULTIPASS. BANK 40 WIDE x 12 HIGH. (See text.)

First Pass				Second Pass				Third Pass				Average Speed Ft. per Sec.
No. of Cols.	Initial Speed	Final Speed	Aver.	No. of Cols.	Initial Speed	Final Speed	Aver.	No. of Cols.	Initial Speed	Final Speed	Aver.	
40	69	0	34.5	..	..	..	..	..	..	..	..	34.5
40	..	..	..	..	..	..	..	..	..	..	..	..
36	77	7.7	42.0	4	69	0	34.5	..	..	..	..	41.2
36	77	7.7	42.0	3	93	23	58.0	1	69	0	34.5	43.0
32	87	17	52.0	8	69	0	34.5	..	..	..	..	48.8
32	87	17	52.0	6	93	23	58.0	2	69	0	34.5	51.8
28	100	30	65.0	12	69	0	34.5	..	..	..	..	55.8
28	100	30	65.0	10	83	14	48.5	2	69	0	34.5	59.3
24	116	46	81.0	16	69	0	34.5	..	..	..	..	62.5
24	116	46	81.0	14	79	10	44.5	2	69	0	34.5	66.0

The matter of proper drainage of condensate from the heating surface of these equipments cannot be exaggerated, for water is a bad conductor of heat, and any small accumulation will materially interfere with the work. This point has been especially well taken care of in the so-called "Criss Cross" design.

With standard vertical tube effects in which the steam is on the outside of the tubes, the design involves different problems, and has gone through a long period of development, until to-day, it can be said that as good results as may be expected are obtained.

Originally, steam was admitted at one point near the top of the cylindrical steam belt as indicated in Fig. 1. It has been established, after many years of experience that under these conditions, it is almost impossible to remove completely the non-condensable gases. Usually, a vent or air outlet is placed at a point diametrically opposite the steam inlet in order thoroughly to purge the apparatus. The tendency of active and quiescent zones of ebullition is to arrange themselves as shown. The

calandria being large in diameter as compared with its height, there is a very large path offered for the flow of steam, rather short as compared with its cross section, and many times the area of the steam inlet. The vapor currents being induced by the venting on the other side, tend to take the shortest path, and the one offering the least friction. Thus, a large portion of the heating surface is shunted, and becomes gas bound, and therefore ineffective. Some designers have attempted to overcome the problem by providing avenues of admission which radiate from the steam inlet, this being accomplished by leaving out certain rows of tubes.

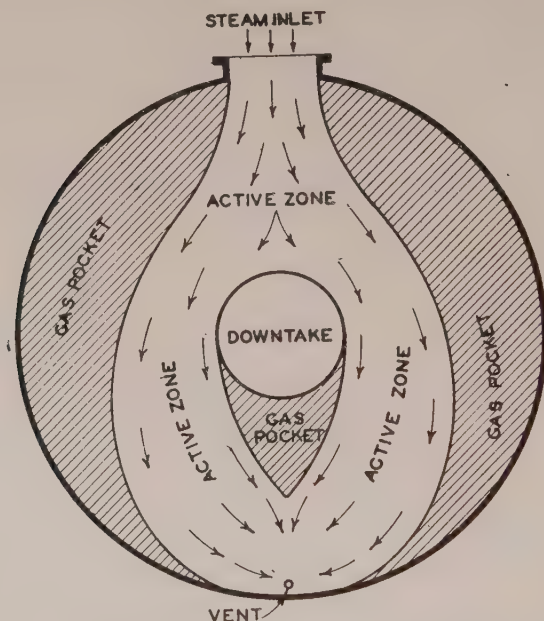


FIG. 1.—Gas Pockets and Vapor Streams in Steam Belts.

This scheme, however, has met with poor success as there is no definite point for logical and positive venting, resulting in a haphazard steam distribution.

Fig. 1 shows the arrangement of active and quiescent zones as actually observed in practice with this design. Such zones are not well defined, but blend gradually one into the other. The location and shape of the gas pockets indicated will vary as the position of the vent is changed, or the operating conditions vary; and, as suggested in another chapter, it is possible materially to improve the work of the evaporator by providing additional taps for gases when their location has been determined by actual observation.

The difficulty explained above led to the design and extensive adoption

of the annular steam belt as shown in Fig. 2. Here, a space is provided all around the calandria, and steam is admitted into this. Slotted holes, or perforations are provided for the entrance of steam into the tube space as indicated. The intention of the designers was to have an appreciable difference of pressure between the annular space and the inside, thus causing steam to flow uniformly through all the openings, thereby giving equal distribution throughout the periphery. In practice, this has been found impossible to accomplish, and particularly so in apparatus of large size.

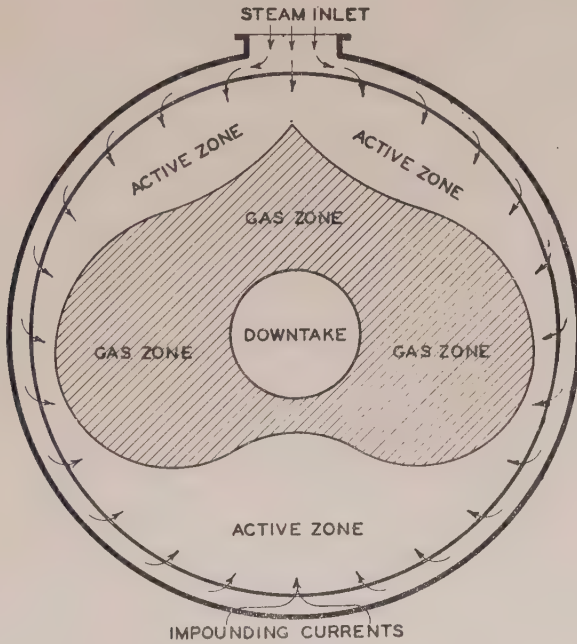


FIG. 2.—Active Zones and Gas Zones in Annular Steam Belts.

The area of the annular space is necessarily limited, and therefore, the flow of steam through it around the steam belt is unavoidably high. This causes a shunting of the openings near the steam inlet, whose flow is divided up into two currents, one to the right, and one to the left. These meet at a point opposite to the entrance, and thus impound one upon the other. The result is that by far the greatest flow is at this very point, as indicated on the sketch which shows the ebullition as actually observed in practice. As in the preceding case, the shape and extent of the gas zone can be altered by changing the position and size of the vents, and the line of demarkation between the gas zones and the steam zones is not sharp and well defined as shown. The sketch has been made up for the purpose of emphasizing the point under consideration.

In very large units, the arrangement shown in Fig. 3 is sometimes used. Here, there are two sets of vapor pipes instead of one. By this means, the flow of vapor is divided up into four streams in the annular space, two proceeding from each inlet, one to the right, and one to the left. In a given sized annular space, therefore, the velocities are halved, which is the same as having twice the cross section. There is another advantage in that there is a double impounding of currents as shown, resulting in a more symmetrical distribution of heat, and in actual operation

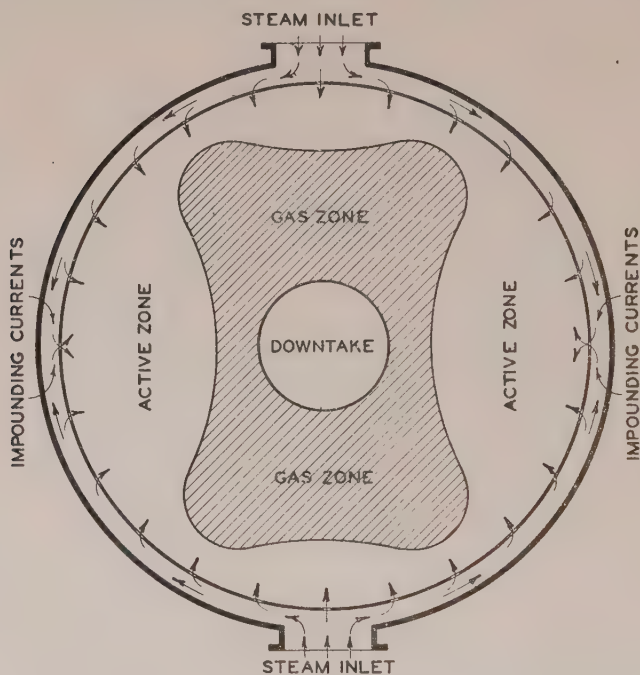


FIG. 3.—Active Zones and Gas Zones in Annular Steam Belts.

it is claimed that better results are obtained. This point is not well established. Undoubtedly, the friction loss in the annular space is much reduced.

Fig. 4 is a slightly different arrangement in which the double inlets are set 90° apart instead of 180° , for the purpose of simplifying the vapor piping. This scheme is probably not as good as the other due to its lack of symmetry, and therefore the probable inequality of the distribution of steam.

There is no limit to the number of vapor pipes for those who insist. There are evaporators with three and even four sets, as shown in Fig. 5, which represents a very popular arrangement in the salt industry. The arrangement of vapor pipes is very complicated and expensive, involving

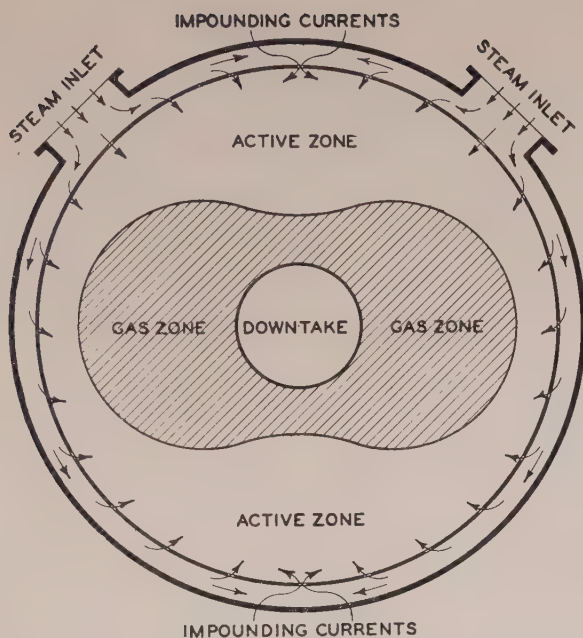


FIG. 4.—Active Zones and Gas Zones in Annular Steam Belts.

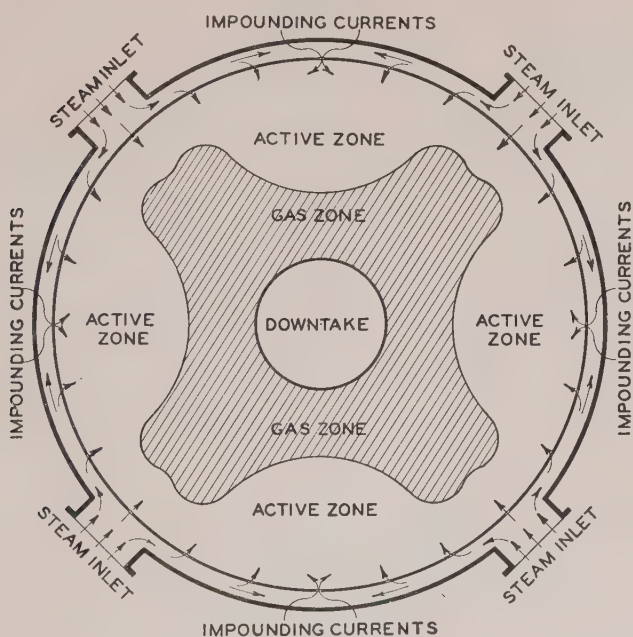


FIG. 5.—Active Zones and Gas Zones in Annular Steam Belts.

many joints and odd shaped castings and fittings. This scheme is used on evaporators of very large diameters from 15 to 30 feet.

Fig. 6 shows the author's idea of a logical annular belt. In this case, the slotted holes or perforations are omitted entirely, there being no partition between the annular space and the tubes. Steam is admitted at two points in such a way as to cause a tangent flow throughout the annular space always in the same direction, and without interruption. Thus it will be impossible for gas pockets to form near the outer periphery, and the surfaces will be subjected to a strong scouring effect particularly towards the outside of the evaporator. The distribution of steam is perfectly uni-

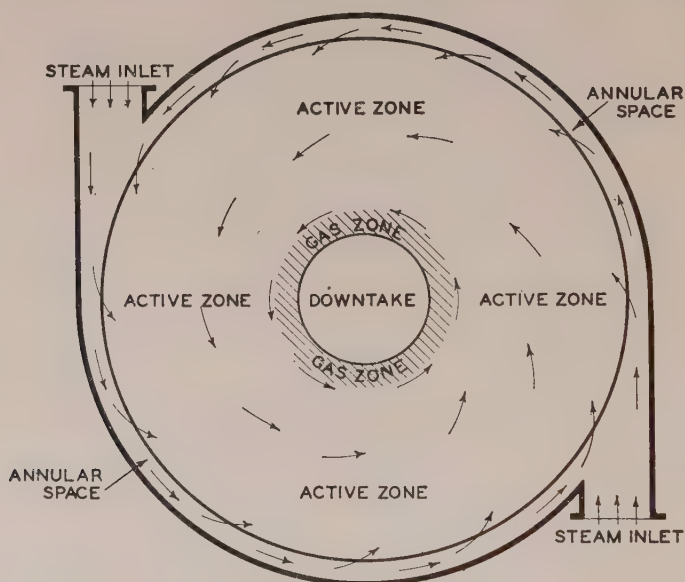


FIG. 6.—Active Zones and Gas Zones in Annular Steam Belts.

form, although the flow of gases towards the vents in the centre is somewhat slow. There should be practically no friction losses after leaving the vapor pipes. This design has recently been used successfully.

In none of the above arrangements is the path of steam through the heating surface well defined, as was mentioned at the beginning of this discussion. The fulfilment of this requirement is accomplished by the use of diaphragms or baffles installed between the tubes.

Fig. 7 shows such an arrangement, which is commonly known as a "zig zag" design. The path of steam through the heating surface is as shown in the sketch, the flow being horizontal, from one side to the other. The main objection here is that on account of the sudden reversals and sharp turns, there are a number of unavoidable pockets of quiescence where non-condensable gases may accumulate. The downtake in the centre

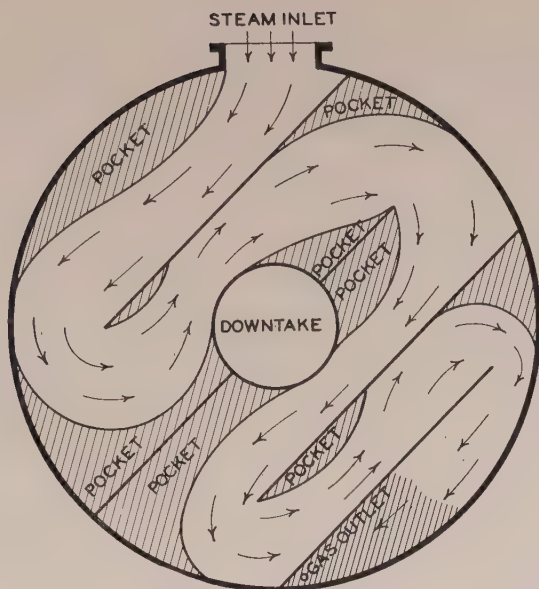


FIG. 7.—Zig-Zig Method of Steam Flow in Steam Belts.

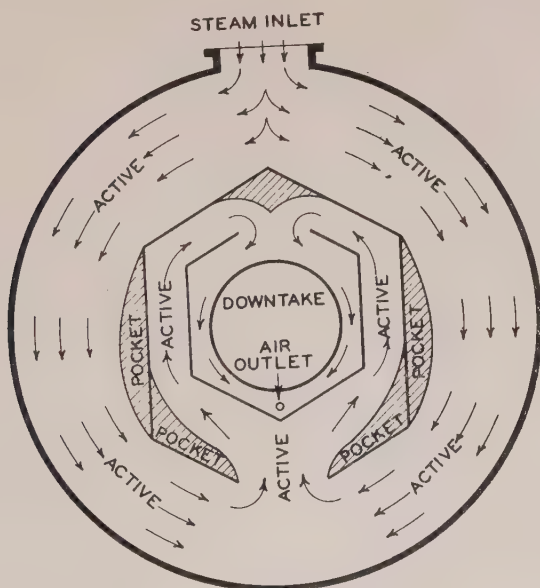


FIG. 8.—Steam Flow Through Webre Baffled Steam Belt.

presents a serious problem, disrupting the continuity and symmetry of the scheme. When multiple downtakes of smaller diameter are used, this objection is overcome to a great extent, but the sudden turns still remain.

Another similar scheme consists of a spiral baffle, winding around the downtake. Here, however, it is very difficult to avoid excessive friction losses, particularly under the conditions prevailing in the last body.

Sketch S.B. 8 shows the writer's arrangement, which has given many years of successful service. There may be slight pockets as indicated, but they are very small, and in actual operation, the surfaces are uniformly active. The design is symmetrical about the downtake. The scheme works equally well with distributed circulation, and the friction losses are very small, practically negligible. The turns are gradual, and the intense zone is always on the outer part, the venting taking place from the centre.

There are also a number of special designs which deserve consideration, and they will be discussed separately under Types of Evaporation in another part of this book.

Chapter 17.

Single Effect Finishing Pans.

In the chapter on the practical use of the heat balance, the matter of the number of bodies in series which can be used for a specific problem was fully discussed.

When the range of densities during operation cover high concentrations and viscosities, involving high boiling temperature losses, the number of bodies in a multiple effect series becomes less and less, and for very high densities, it is sometimes possible to operate only one body within the available temperature range as defined by the pressure of steam on one hand, and the maximum obtainable vacuum on the other. This results in single effect operation, so called.

When the percentage of evaporation by weight from the initial to the final density is very great, it sometimes pays to split the work into two parts, the major, or low boiling range being done in multiple effect, with resultant steam economy, and the smaller or high boiling range being done in a single effect, utilizing the full temperature fall on this body, which, in this case, we call a finishing pan. In this way, it is possible to do most of the work in a set having a number of bodies larger than could be used if the entire concentration were attempted in multiple.

Thus, in the sugar industry, the evaporation of juice into syrup from say 15 per cent. solids to 60 per cent. solids is done in multiple effect, and the final concentration to crystallization is done in a vacuum pan, which is a single effect finishing pan. It is interesting to note that the first multiple effects installed by Rillieux in the sugar industry were triple effects, in which the whole concentration was carried on, the last body acting as the vacuum pan, and the process being intermittent as far as the evaporating station was concerned.

Similarly, in the manufacture of tanning extract, where the percentage of evaporation by weight runs as high as 95 per cent. or more, it pays to do 90 per cent. of this work in multiple effect, and the balance in a finisher. The exact point at which this will pay depends on local conditions and the ultimate density to which the concentration must be carried. When the final density becomes too great, however, the temperature drop required to operate the last body becomes so large, that the entire equipment is slowed down. Furthermore for such conditions, the tube proportions must be materially altered giving an equipment not adapted to the work, involving disproportionate sizes, and even a decrease in the number of bodies in series, thus vitiating the cycle.

The circulation of liquor at high densities is a matter of prime impor-

tance, and the finishing pans must be designed for fast liquor circulation and slow vapor velocities by the use of tubes of small lengths and large diameters. It may be advisable to supplement by mechanical circulation in extreme cases.

There is nothing extraordinary about a finishing pan. It is merely a single effect evaporator designed to operate on viscous heavy liquors. It is usually operated by the "batch" system, in which the contents of the evaporator are concentrated "dead end" up to the finishing point, after which the batch is dumped, and a new one started. In this way, better control of the ultimate density is obtained, as the contents of the apparatus are cooked up to the prescribed point. It is also true that when operating on the batch system the heating surface is subject to the impeding action of the final density only towards the last part of the strike, being free to operate at a fairly high speed at the beginning, when the density is comparatively low, as it is received from the multiple effect preceding. If concentration is by the continuous method, where partially concentrated liquor from the multiple is fed directly into the finisher, and pumped out continuously as finished product, it is very evident that the heating surface of this unit must suffer at all times from the ultimate density, and therefore its operation must be necessarily slow. On the other hand, there is an advantage in that there is no interruption caused by batching. Continuous operation is advised where possible.

Some finishers are arranged so that they can operate continuously only, as they have no storage for the batch. The other type is much to be preferred as it is easier to control, particularly by the ordinary plant operator.

If the desired ultimate concentration is very high, and involves particularly viscous materials, it may become necessary to use live steam even though the operation is under vacuum. Physical characteristics will determine the exact method of operation.

It is well to call attention to the fact that the high viscosity of the finished products under consideration is an inverse function of their temperature, and it is sometimes advisable to operate the evaporator with relatively low vacuum, in order to reduce the viscosity by raising the boiling temperature. Certain organic materials are injured by high temperatures, and thus require high concentration under high vacuum, in which case the expedient referred to above cannot be utilized.

Finishing pans designed for fast liquor circulation should have ample downtime area to provide for the easy return of the liquor circulated from the top to the bottom of the apparatus.

In providing the separator between the pan and the condenser, it must be designed with ample drains, sealed in the liquor below. Such drains have a tendency to plug up, and should be inspected at frequent intervals. Many vacuum pans in the sugar industry have been found upon inspection to have the drains from their catchalls completely plugged up with hard sugar, under which conditions the losses by entrainment must have been very considerable, as the work of such a catchall is nullified.

It should be noted that entrainment losses at the finisher are exceptionally important, the product being highly concentrated and consequently much more valuable, and losses much more serious.

It is true that the fouling of heating surfaces at extreme concentrations is much less than for lower densities, such as occur in the multiple effect.

Some operations for ultimate concentration are done with direct fire in order to keep up the high temperatures during this period. Such is the case in finishing caustic soda, calcium chloride, etc. With both of the above, complete solidification takes place on moderate cooling, as even the water of crystallization has been driven off.

Chapter 18.

Evaporator Condensers.

The general problems involved in the condensation of vapors from the last body of a multiple effect evaporator are practically the same as those encountered in the condensation of steam from a turbine, with the one difference that the quantities of air and non-condensable gas to be removed from the condenser are usually very much greater. For this and other reasons, it does not pay, and it is not advisable to carry very high vacua except under special conditions.

It will be well to review the functions of condensers, their general design, and the problems involved in their operation. This discussion will be confined to the subject of jet condensers, which are in almost universal use in evaporator work.

There are two well-defined functions of condensers while maintaining the desired vacuum, namely: first, to heat the water introduced to as near the temperature of the entering steam as possible; second, and almost as important, to cool the air and non-condensable gases, which are normally removed by the vacuum pump, to as low a temperature as possible. The first of these functions makes for economy of water, and the second, for economy of vacuum pump displacement.

A condenser can be called 100 per cent. efficient if the injection water leaves the tail pipe at the same temperature as that of the vapor entering, and if the air and non-condensable gas removed by the vacuum pump are maintained at the same temperature as the entering injection water. Surely, this would represent perfect operation and in our experience we have never seen it accomplished.

Some manufacturers make statements for advertising purposes which are open to serious misinterpretation by giving the vacuum produced by their condensers as representing the percentage of obtainable vacuum, based on the temperature of outgoing injection water. The evaporator engineer is interested in the temperature range of his evaporator, which determines its capacity, and in order to obtain a true valuation of the work of the condenser, it is necessary to base all the information on temperatures rather than pressures. This is also true if the condenser is applied to apparatus other than evaporators.

The efficiencies calculated by the two methods are very much at variance, as is shown in the following example:

Assume a condenser operating in conjunction with a spray pond in

the tropics, and that fair average conditions are represented by the following:

Injection water entering condenser.....	90° F.
“ “ leaving “	110° F.
Vacuum obtained	26 in.
Air leaving condenser, going to vacuum pump.....	110° F.

Estimating the performance of the condenser on the basis of temperatures as suggested above, there would be:

Temperature corresponding to vacuum.....	125° F.
Entering water	90° F.
Maximum possible range of heating.....	35° F.
Temperature of water leaving condenser.....	110° F.
“ “ “ entering “	90° F.
Actual range of heating.....	20° F.

The water efficiency of the condenser is the ratio of the actual heating range to the possible heating range, or in this case, $\frac{20}{35} = 57.2$ per cent.

The efficiency of the condenser with reference to air cooling would be as follows:

Vapors and gases entering the condenser.....	125° F.
Temperature of same leaving condenser.....	110° F.
Range of cooling.....	15° F.
Vapors and gases entering the condenser.....	125° F.
Temperature of injection water entering.....	90° F.
Theoretical possible range of cooling.....	35° F.

The air cooling efficiency is the ratio of the actual cooling to the possible range, or $\frac{15}{35} = 43$ per cent.

The calculated efficiency by the other method is radically different, and very much higher, as is shown below:

Vacuum corresponding to outgoing water temperature. Vapor pressure corresponding to 110° F. = 2.59 in. hg.,

Assume 30 in. barometer, vacuum = (30 — 2.59 = 27.41).....	27.41
Vacuum actually obtained.....	26.00

The efficiency of the condenser according to this method is the ratio of the above two figures, or $\frac{26.00}{27.41} = 94.8$ per cent.

The discrepancy between the two methods of calculation will vary according to the performance and conditions, but it is always disproportionate

and misleading, and the vacuum ratio should therefore not be used to indicate efficiency in this work.

The proper cooling of air and non-condensable gas in evaporator condensers is of greater importance than it would be in ordinary condenser practice, inasmuch as there is a relatively much larger volume to be considered, and the bad effects on the operation of the condenser are therefore much more serious.

In the illustration given above, it must be borne in mind that there are non-condensable gases entering the system from two sources. First, there is the normal amount of air dissolved in the injection water, in quantity usually determined by the temperature of this water. (See table in appendix for saturation of air in water at different temperatures.) In addition to this there may be other non-condensable gases in solution, depending upon the source of this water supply, which must all be taken into account. There also enters into the condenser, with the vapor to be condensed, air from vacuum leaks in various parts of the apparatus operating under vacuum. These vacuum leaks may be quite considerable, for the amount of air which enters through a small orifice is quite surprising. In fact, to the layman, it is almost unbelievable. The volume thus admitted will depend upon the care and attention with which the vacuum joints have been made, and will vary greatly with different installations. The air enters through these leaks dry and at atmospheric pressure, and reaches the vacuum pump heated, vapor saturated, and expanded by the reduced pressure, resulting in a relatively large volume.

The sum total of this air and vapor must be removed by the vacuum pump. The volumetric capacity of the pump as a rule is fixed, and always determined beforehand when the apparatus is purchased. The calculations for size are based upon certain assumptions supposed to represent the expected operation under probable fixed conditions. It generally happens that this operation is not exactly as expected, and it always happens that the conditions are not fixed, but on the contrary quite variable. The result is that for any particular vacuum in the condenser, there is an exact volumetric vacuum pump displacement, no more, and no less. Incidentally, it might be mentioned here that the actual displacement of this pump as compared with its swept volume varies considerably with the vacuum maintained, and decreases rapidly as the vacuum is increased. Its displacement, however, is fixed for any given set of conditions. If the quantity of air and accompanying vapor in operation is not enough to supply the pump under estimated vapor pressure relations, then, of necessity, the ratio of vapor to air will increase, swelling the volume of the mixture to satisfy the requirements. The converse is also true, namely, if the volume of air is more than the estimated amount, the ratio of vapor to air will decrease until the volume of the mixture corresponds to the displacement of the pump.

These conditions have a very important bearing upon the operation of the condenser, for they decrease or increase the amount of non-condensable gas in transit in the zone of condensation, improving or depreciating the transmission of heat from vapor to water therein.

It is readily seen, therefore, that the air cooling efficiency of the condenser will fluctuate with the displacement of the vacuum pump, and that normally, inasmuch as the water surface exposed in the condenser is limited, as the vacuum pump displacement increases, the water heating efficiency will increase, and the air cooling efficiency will decrease. The converse also holds good, and it is found in practice that a selection must be made governed by existing conditions to give the most desirable result.

Water is not a good conductor of heat, and therefore, the condenser which breaks up the water into relatively thin sheets or small drops is better than the one in which the injection passes through in thick, slow moving, bulky volumes.

The presence of air, even in small proportions, in the zone of intense condensation materially interferes with the absorption of vapor and its accompanying heat by water. For successful operation, it is essential to remove this air as rapidly as possible from these parts, as its accumulation brings about poor results by either lowering the vacuum or increasing the water consumption to maintain the same vacuum. In fact, where conditions are difficult, it might be better to use a supplemental air cooler, using a separate air chamber on the outside, with a fine spray of cold injection water. This is probably not advisable in small installations.

Referring again to the specific illustration at the beginning of this discussion, assume the following conditions:

Vacuum in condenser.....	26.00 in.
Corresponding absolute pressure.....	4.00 in.
Temperature of air leaving condenser.....	110° F.
Corresponding vapor pressure.....	2.59 in.

The mixture of air and vapor leaves for the vacuum pump at an absolute pressure of 4.00 in. of mercury. This represents the total pressure. The partial vapor pressure corresponding to the temperature of this mixture is 2.59 in. of mercury, and the air pressure is the difference between the two, or 1.41 in. of mercury, so that all of the non-condensable gases must leave the condenser at this latter pressure for removal, or at a vacuum of 28.59 in., based on a 30-in. barometer.

With a well designed air cooler, it should be possible to cool this mixture to the temperature of the incoming water, or in this case, to 90° F. The vapor pressure corresponding to this temperature is 1.417 in., and the air pressure would therefore be 4.000 in. — 1.417 in. = 2.583 in., which in turn corresponds to a vacuum of 27.417. This means that the pump displacement need only be about 55 per cent. of what it would have been in the other case. The air cooling efficiency is thus increased greatly.

By increasing the volume of water injected into this cooler, the ratio of vapor pressure to air pressure at the outlet of the condenser can be increased to any desired proportion within reason, and this would naturally bring about a higher water heating efficiency, inasmuch as the interfering air pressure in the zone of condensation would be practically eliminated.

As a matter of fact, condensation and air cooling are substantially separate physical reactions, requiring separate provisions, and it is possible

to handle them better separately in apparatus designed for the purpose, rather than make a compromise in one apparatus that does neither function well.

Another signal advantage of such a combination is that very often, in a given plant, besides the supply of condensing water which may be relatively warm, there is another smaller supply of very much colder water which might be used in the cooler suggested above, bringing the air down to such a temperature that the partial vapor pressure would assume almost negligible proportions, at the same time displacing such a large volume from the condenser as to permit vapor heating of the injection water to almost vacuum temperature.

Types of Condensers. There are, generally speaking, three types of condensers in common use for evaporator work as follows:

1. The counter current condenser, in which the vapor to be condensed enters the lower part, and the air leaves the upper part. The water enters the top and drops down, flowing in the opposite direction to the vapor, hence the name.

2. The parallel current condenser, in which both water and vapor enter the upper part, and leave at the bottom.

3. The ejector condenser, in which both water and vapor enter the top and leave the bottom, the lower part of the condenser being so designed that the mechanical energy of the falling water carries with it the non-condensable gas which has thus been trapped out.

The characteristics of each will be discussed briefly.

The Counter Current Condenser. The counter current condenser is generally accepted as representing the best practice to-day, and the arguments in its favor are well taken. From the very nature of its operation, one would conclude that everything else being equal, it will heat water more nearly to vapor temperature, and cool air more nearly to water temperature. There are a number of variations in design, the best being, as stated before, the one in which the water is more finely divided.

Certain condensers have special characteristics which it is well to discuss. One is the difference of pressure between the upper and lower part of the condenser. Especially in the old types, where the downward motion of the water is very slow, and arrested at several points almost completely by a succession of trays and cascades, resulting in bulky masses of slow moving water, it very often happens that there is a considerable loss of pressure between the bottom and the top of the condenser, such losses having been observed at as high as 1 in. of mercury, or even more.

Usually, the water flows downwards in circular sheets, and in making the deflections or steps, it is best not to completely arrest the flow by the use of flat plates or trays, but rather to accomplish the breaking up of the streams by means of conical surfaces with fairly steep angles, thereby permitting a continuous acceleration of the downward movement, resulting in finer subdivision of water, more surface, and better exposure.

It is advisable to provide openings in these sheets of falling water for the upward passage of the air and non-condensable gases. In this way, two objectionable features of the condenser are overcome. In the

first place, there is practically no difference of pressure between the top and the bottom. In the second place, it is not necessary for the gases to rupture the sheets of falling water in order to reach the air outlet at the top of the condenser. Without such provisions, this counter current con-

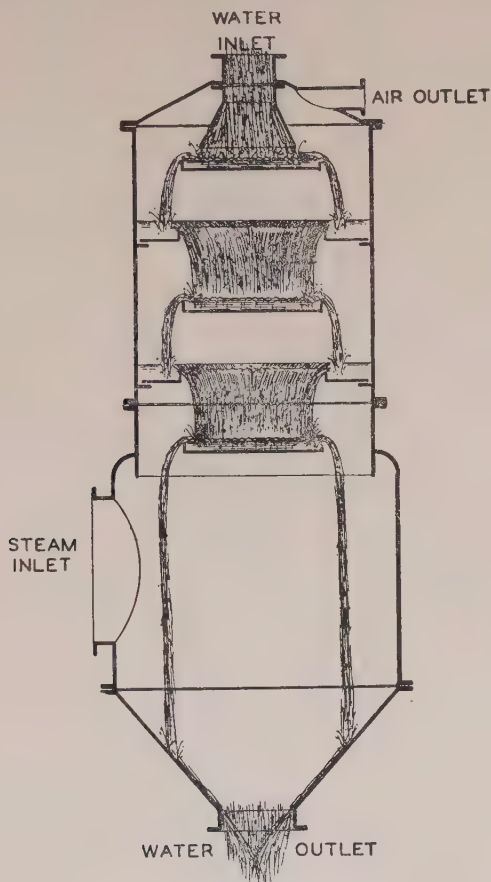


FIG. 1.—Old Style Cascade Counter Current Condenser.

denser is given to "boiling over," or frothing of water into the air outlet, a bad feature, which is aggravated where the vacuum pump displacement is relatively large, or where a considerable volume of non-condensable gases have to be removed.

Provisions are generally made against boiling over by installing a separator in the line to the vacuum pump. This provision is specially important if the system is provided with a high speed dry air pump having small clearances, as any sudden material amount of water picked up from the

condenser and reaching the pump will smash it. Many vacuum pumps have been wrecked by neglect of this precaution.

The counter current condensers when well designed are quite satisfac-

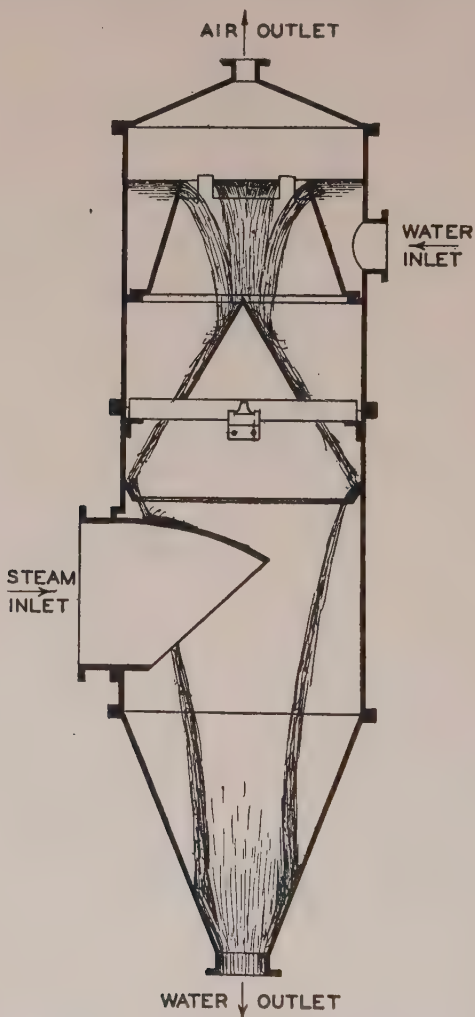


FIG. 2.—Weiss Type Counter Current Condenser.

tory in operation, easily controlled, and give good results. They are flexible as to load, variations of vacuum, and variations of water temperatures. In Cuba, their use is almost universal. Sometimes, they are built in very large sizes with several vapor inlets for various pans and evapora-

tors, each being provided with a large valve, so that one unit can be shut down without interfering with the other.

Parallel Current Condensers. Where the initial temperature of the injection water is relatively cold, there is little doubt that it is practically impossible to better the performance of a well-designed counter current condenser. If, however, as is generally the case in warm climates, the range of heating of the water is determined by the cooling of the spray pond, or cooling tower, there is not much to choose between a counter

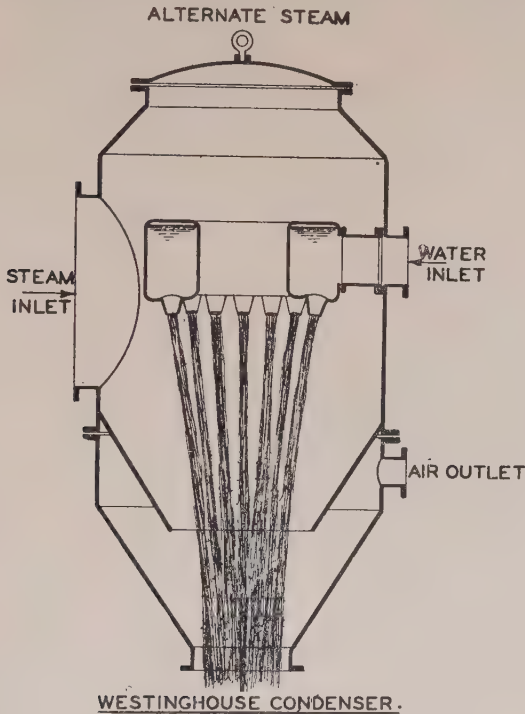


Fig. 3.—Excellent Example of Good Parallel Current Design.

current and a parallel current condenser, both being equally well designed, and provided with suitable pumps.

The cooling range of a spray pond under good operating conditions is about 20° F., and that of the average cooling tower, such as is found in the tropics, is somewhat less, depending upon the wet and dry bulb temperature of the air, and the wind conditions.

The modern parallel current condenser is equipped with spray nozzles for water admission, and consequently exposes a tremendous amount of surface, probably considerably greater than the ordinary counter current condenser. The initial cost is relatively considerably less than the other,

and the net results are about the same. A careful survey of the comparative performance of these condensers in Cuba conducted some years ago revealed that there is little difference in their operation. No doubt, this is not true for any condenser problem, but it apparently holds good for tropical conditions, where the range of heating is small.

There are parallel current condensers available to-day which will heat the water within 10° of the vapor temperature, the non-condensable gases, of course, leave at the same temperature as the outgoing water, but the required vacuum pump displacement is reduced due to the fact that a large part of these gases is carried away in the tail pipe by the falling water.

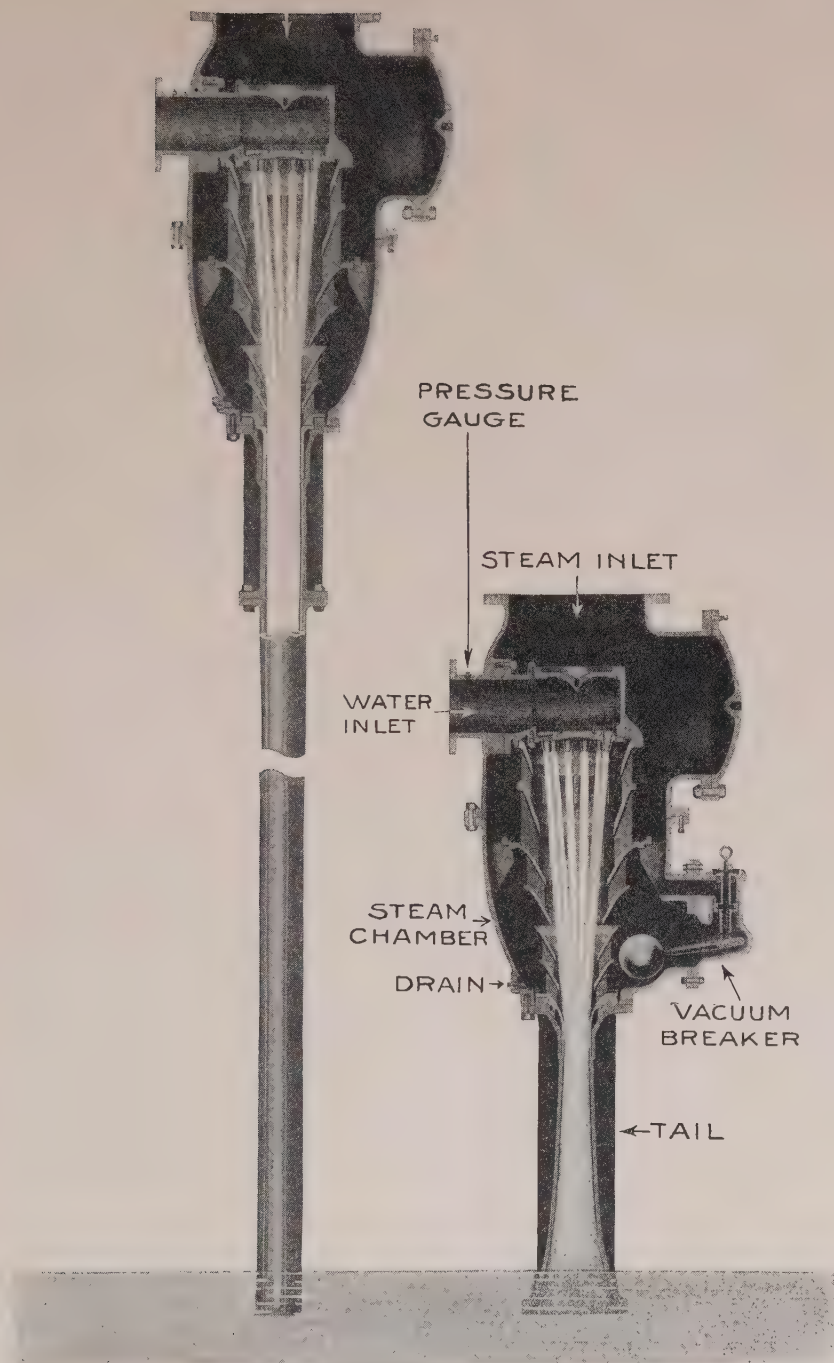
In counter current condensers, operating under the same conditions, the outgoing water and air would be about the same as for the above, under the same temperature and vacuum relations.

If the range of heating increases very materially, then the advantage quickly turns in favor of the counter current apparatus, and the divergence is greater as this range increases, for the temperature of the outgoing air will then be much colder if it comes in contact with the colder water at the top of the condenser, instead of being mixed with the hot water below.

Ejector Condensers. Until recently, the ejector condenser has always been looked upon with disfavor, the main reason being its reputation for excessive water consumption, its lack of flexibility, and the usual prejudice against deviating from precedent. In a number of recent installations, however, its performance has been so satisfactory, that it can no longer be denied that it has a useful field of application, and where favorable circumstances permit must be given consideration.

In the first place, such a condenser, from its very nature, will not eject unless water is supplied to it in sufficient quantities and under sufficient pressure that the rapidly moving streams from the nozzles will completely fill the combining throat at the bottom, the kinetic energy at this point being sufficient to overcome atmospheric pressure, and in excess, being able to drive out with it entrapped bubbles of air contained either in the original water, or otherwise entering the system. The installations to which reference has been made above were equipped with ample barometric columns, although the manufacturers advised that this was not necessary, the only variation involved being a readjustment of water pressures to compensate for the increased pressure at the combining throat.

Be that as it may, observations made on installations with barometric columns show that it is easily possible to obtain vacua up to 27 in. with water at 70° F., even when there is no pressure at the water inlet. In fact, this was actually done with a vacuum at this point of 12 in. to 15 in. This particular unit had been designed to operate with 0 pressure at the water nozzles. This performance shows that there is quite a material flexibility for the water rate, and places the apparatus in a class where it must be seriously considered. The measured air handling capacity showed also that it was amply able to take care of itself from this score, a feature of particular interest for evaporator work.



Courtesy Schutle & Koertnig Co.

FIG. 4.—Ejector Condenser.

The most important asset, however, is the complete elimination of the vacuum pump, with its attendant moving parts, upkeep, and lubrication, resulting in a very material simplification, so that even if the water consumption is somewhat greater, it must be credited with its other merits.

In large units, the nozzles will pass obstructions up to 1 in., and in practice, very little trouble from clogging of these has been found.

One successful installation in the tropics is operated in connection with a spray pond. The vacuum maintained by the ejector is as good as that maintained by other condensers doing the same work at the same plant.

It is possible to make quite a simple and satisfactory combination of the ejector condenser with a spray pond by installing it at an elevation so that the hot well water will flow by gravity to the pond. The pumping is then done in one operation, from the pond to the condenser. Thus there is only one water pump, instead of two water pumps and one vacuum pump, as at present in many plants.

The barometric installation is preferable to the low type setting, and probably gives more reliable work, and doubtless has a larger air handling capacity.

Chapter 19.

Spray Ponds and Cooling Towers.

The capacity of an evaporator is limited by the amount and temperature of the condenser water available. Those plants which are fortunately situated in this respect have a considerable advantage over others in that abundant cold water permits greater temperature differences between the initial steam temperature and condenser temperature, and therefore the possibility of a greater number of effects with the corresponding saving on steam consumption.

Where condenser water is insufficient in quantity, it is possible to use it over again by cooling it after use by bringing it into contact with the air.

Air which is unsaturated has the property of causing water to evaporate until it becomes saturated if left in contact with water a sufficiently long time. This evaporation requires heat, and therefore, the water may cool, or the air may cool, or both may cool at the same time, depending upon conditions. This property of air is made use of for the cooling of condenser water.

There are three commercial methods for cooling water with air, the cooling pond, the spray pond, and the cooling tower. These differ only in the method of bringing the water and the air into contact with each other.

The cooling pond consists of a small pond or reservoir into one end of which the hot condenser water is pumped, and out of the other end of which the cool water is sent back to the condensers. The air blows over the surface of the pond cooling the water by quiescent evaporation. The cooling capacity of these ponds is small per square foot of surface area, and the cost is therefore large on account of the large size required. This type of cooling is used only where ground space is inexpensive, and there is a good natural bottom to prevent loss of water by natural seepage. There will be a certain loss of water in all cases due to evaporation, which must be made up by the addition of make up water from some other source. This amount is small in this type of pond, and should not exceed 2 per cent. of the water cooled, except in the case of very high temperatures.

The capacity of cooling ponds depends entirely on the atmospheric conditions at the particular locality in question. A rough figure used by engineers for this type is 3 B.t.u.'s per hour per square foot of pond surface per degree F. temperature difference between the water and the air. Air that has a high average humidity will not cool as quickly as this, while air that is very dry will cool considerably faster.

It is usually more economical to reduce the pond area required for

cooling a given amount of condenser water by the use of sprays, in the well-known spray pond. The chief difference in the two types of pond is in the capacity per square foot of pond surface. The cost of the sprays and piping and the cost of pumping the water amount to a considerable item which must be considered as well. In windy climates, the loss of water by entrainment in the wind may increase the amount of make-up water required considerably above that required in the cooling pond.

The capacity of spray ponds is much greater per square foot of pond surface than is the case with the cooling ponds. Using the figure for the



FIG. 1.—Spray Cooling Pond.

heat transfer for cooling ponds given above, under average conditions as to air temperature and humidity, a cooling pond may be expected to cool about 5 pounds of water per hour from 100° F. to 70° F., while a modern spray pond should cool at least 150 pounds of water per hour per square foot of pond area.

It should be noted that in both types of ponds, it is impossible to cool the water below the wet bulb temperature of the air. In dry regions, where the humidity is low, there is a large difference between the temperature of the air and that of the wet bulb, and under these conditions, it is possible to cool the water to well below the temperature of the air. On the other hand, in very humid regions, this is very difficult.

Insofar as present practice is concerned, the advantage is apparently

on the side of the spray pond, as the authors do not know of any recent installations in their vicinity where cooling ponds have been put in. A typical spray pond is shown in Fig. 1.

Where ground space is too valuable or is restricted, it is possible to make use of cooling towers, which are equivalent to cooling ponds placed on end.

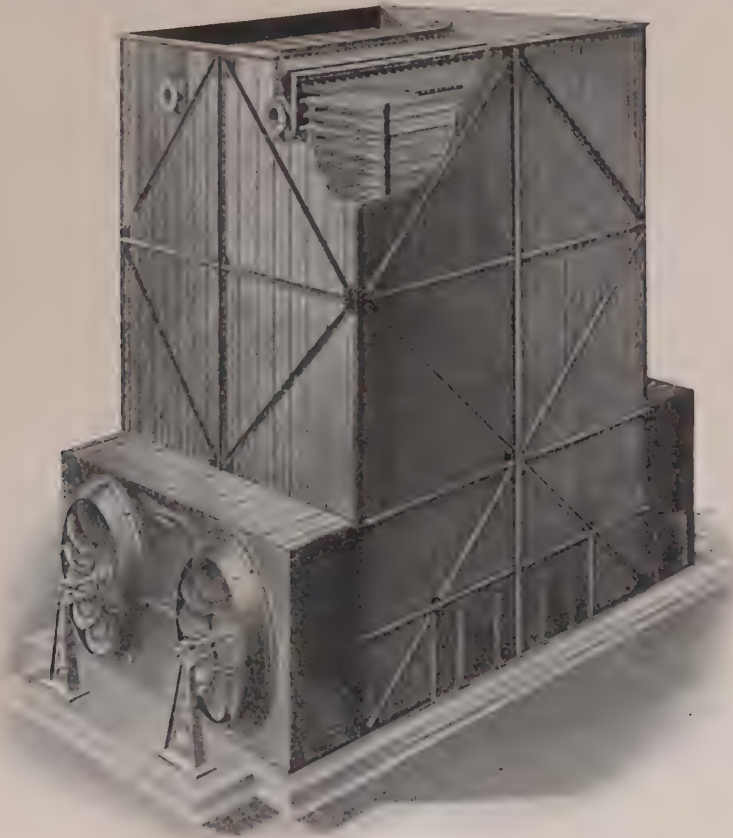


FIG. 2.—Forced Draft Cooling Tower.

There are three classes of cooling towers, namely, atmospheric, natural draft, and forced draft. The names are quite descriptive. The natural draft tower consists of a vertical shell surmounted by a chimney. The tower is filled with slats or other variety of lattice work, over which the hot water is allowed to trickle. The air enters at the base of the tower on account of the difference between the weight of the column of warm air inside the tower and chimney and the weight of the cooler air outside,

the air is forced up through the tower in counter current flow to the descending water, thereby cooling it to a temperature in the vicinity of the wet bulb temperature of the entering air. The efficiency of the tower is measured in terms of the degree of nearness of approach of the temperature of the water leaving the tower to that of the wet bulb of the entering air. It seems obvious that a natural draft tower will not

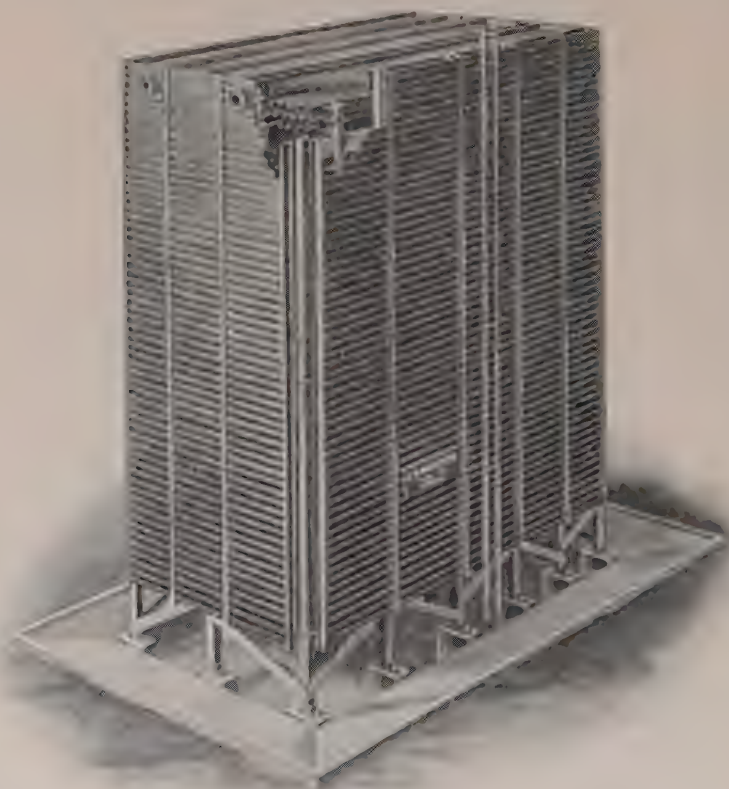


FIG. 3.—Atmospheric Cooling Tower.

operate well unless the average temperature of the air within the tower is considerably above that of the air outside. In other words, such a tower is not to be used where it is attempted to cool the water much below that of the air.

The forced draft cooling tower depends upon the use of a fan or blower, or of some type of injector to force the air through the tower. On this account it is possible to cool the water to a lower temperature than is the case with a natural draft tower, but this is done at the cost

of power required to operate the fan or blower. Furthermore, the fan may become covered with ice in severe winter weather, and cease to operate. The capacity of the tower per cubic foot of volume is greater, however, on account of the increased air velocity possible. A characteristic tower is shown in Fig. 2.

The commonest type of cooling tower is the atmospheric type illustrated in Fig. 3.

This type of tower is open on the sides so that the wind may blow through it and cool the water as it falls down over the lattice work. It has the advantage that it permits temperatures of the cooled water nearly down to wet bulb, but has the disadvantage that its capacity depends to a considerable degree on the amount of wind that is blowing.

The capacity of cooling towers varies greatly with the type of tower and local conditions, but it is customary to use about one square foot of ground area of forced draft towers per 6.5 gallons of water cooled per minute, while the atmospheric towers will cool from 1 to 2 gallons of water per square foot of ground area per minute. The towers range in height from about 20 feet for forced draft, to 60 to 80 feet for natural draft installations.

The loss of water due to combined evaporation and entrainment in such arrangements is about the same as that of spray ponds, that is in the vicinity of 3 per cent. of the water delivered. In the case of spray ponds, where the wind is high, and where the loss of entrained water is serious, it is usually customary to install louvres or fences around the pond to catch as much of the drift as possible, both for the purpose of preventing a nuisance and to conserve water. The authors know of industrial plants in this country using evaporative equipment, where the scarcity of condenser water is so great that it is difficult to obtain even the 3 per cent. of make up water necessary to operate the spray pond.

Chapter 20.

Multicurrent Solution Heaters.

The problem involved in the heating of solutions is parallel to that encountered in the operation of surface condensers, and under ordinary conditions, consists of condensing steam on one side of the heating surface by absorbing the heat so released in raising the temperature of the solution which is being circulated on the other side.

The apparent simplicity of the problem has led to many misconceptions, and the large number of variables have not always been taken care of by the investigators, resulting in wide discrepancies and lack of agreement in the reported results. Below is given a list of these affecting the final result.

Proportions of Tubes.

1. Thickness of metal wall;
2. Diameter of tubes;
3. Length of tubes;
4. Material from which tubes are made.

Physical Conditions.

1. Temperature of steam;
2. Initial temperature of solution;
3. Final temperature of solution;
4. Speed at which solution is passed through;
5. Speed at which steam travels past heating surface;
6. Method of removing condensate;
7. Method of removing non-condensable gas;
8. Cleanliness of heating surface;
9. Character of liquid flow, as to turbulence or lack of same;
10. Power cost of producing circulation;
11. Physical characteristics of solution.

In studying this problem, it is necessary to consider all the facts and their bearing upon the final results. Nothing should be guessed, and it must be remembered that it is dangerous to deduce general conclusions from small scale operations carried on for short periods of time. Nor should such conclusions be drawn from a few isolated facts. Important data should be thoroughly confirmed before placing too much reliance on their dependability.

It has been found that in making analyses of heat transmissions, it is better to consider the resistance to the passage of heat as made up of three parts as follows:

A. First, the resistance to be overcome in making the heat enter the metal of the tubes, commonly called the steam film resistance;

B. Second, the resistance to be overcome in causing the heat to cross the metal of the tubes;

C. Third, the resistance to be overcome in order to compel the passage of heat from the metal of the tubes into the liquid being heated, known as the liquid film resistance.

Extensive and careful experiments have fairly well established the transmission of heat through the steam film, and under ideal conditions, this has been found to be from 2000 to 4000 B.t.u.'s per square foot per hour per degree drop, according to the speed of the steam traveling over the heating surface. This transmission is also affected by the presence of non-condensable gas, which acts in the nature of an insulant, and must be particularly guarded against when the steam pressure is below the atmosphere, as such conditions make non-condensables difficult to remove and larger in volume. The necessary precaution is rapid travel of steam, controlled by compelling it to pass through a predetermined path, preferably of progressively reducing cross section, so as to compensate for volume reduction brought about by condensation, as it proceeds through the heating surface from the beginning to the end of its travel. Thus it is easily possible to maintain fairly high velocities throughout without undue friction losses.

Proper arrangements must be made for the thorough removal of the water of condensation by the use of rain plates which catch the drip, thus preventing it from flooding the lower tubes. Neglect of this detail materially increases the steam film resistance by forcing the heat to travel through a superposed layer of water, which is known to be a poor conductor of heat.

The rate of heat transfer through the metal itself has also been carefully determined, and will be found tabulated in Chapter 3 for various materials. The resistance here is proportional to the thickness of the metal. Usually, however, this resistance is the smallest of the three, so small, in fact, that where due to the nature of the case, relatively low transmissions are expected, it can almost be left out of account.

When liquid is flowing through a tube on the outside of which steam is being condensed at a temperature higher than that of the liquid, heat must enter this liquid from the outer skin of the moving stream which is in contact with the metal surface. As soon as this skin has been heated, further heat absorption can only be accomplished by additional penetration through an ever increasing layer of already partially heated solution, until finally the end of the tube is reached. Since the heat conductivity of the solution is poor, one may expect a rather rapid depreciation of the coefficient of heat transmission as the thickness of the liquor film to be traversed by heat increases during the passage of the liquid through the tube from one end to the other.

Unless the flow of liquid is of extreme turbulence, a condition never encountered unless artificially produced, its velocity at the centre of the tube will be greater than at the outer periphery, and the core or centre part of the stream will have a tendency to by-pass with little heating. This fact is well established by actual observations with thermometers, and it is more or less true as the diameter of the tubes is greater or smaller. Even if the flow were uniform throughout the diameter, it must be ad-

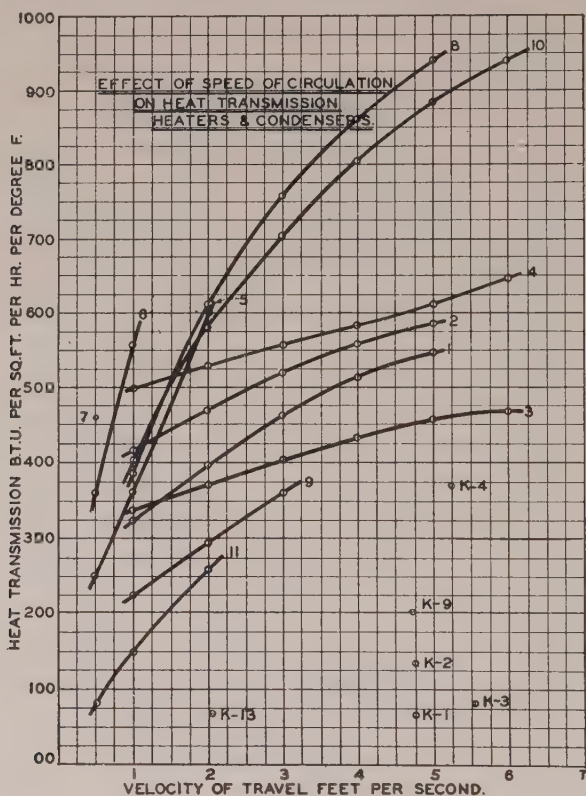


FIG. 1.

mitted that heat penetration would become more and more difficult as the tube diameter increases.

The logical conclusion is, therefore, that the best transmissions are obtainable with tubes of small diameters and short lengths, which again brings out the danger of deducing general conclusions from experimental work on small scale apparatus whose tube proportions differ radically from those of full-sized apparatus.

Actually, the diameters and lengths of heater tubes are determined by purely practical considerations, such as tendency to foul and plug,

mechanical weakness, ease of cleaning, facility of repairs, etc., so that commercial heaters are usually provided with tubes whose diameters vary from $1\frac{1}{4}$ to 4 inches, according to conditions of operation and design, and the relations between their performance and that of experimental apparatus having $\frac{3}{8}$, $\frac{1}{2}$ and $\frac{5}{8}$ inch tubes is problematical. Many observations of heaters under actual operation leads to the conclusion that better results are obtainable from short tubes, for the film penetration is not so serious an impediment.

In practice, the liquor film is broken up to a great extent by compelling the liquid being heated to pass through a number of tubes in series, sometimes as many as 18 in one apparatus, so that the heat stratifications are broken up at the end of each pass, thus producing a homogeneous mixture of heated and unheated liquid at the entrance of each pass. In this way, it is quite apparent that the heating surface at the inlet end of the tubes does considerably more work than the balance, as the film begins here, and increases progressively being maximum at the discharge end.

By far, however, the greatest obstacle encountered in the transmission of heat is the accumulation of scale and dirt on the liquor side of the tube, which in the average case, greatly exceeds the sum of all the other resistances put together. This necessitates the consideration of purely local conditions of operation which are usually elusive and uncertain, and may be subject to great variations even in the same plant. Liberal allowances must therefore be made for contingencies where practice has not been well established.

HEAT TRANSMISSIONS IN CONDENSERS AND HEATERS FOR VARIOUS SPEEDS OF CIRCULATION.

Authority	Tubes	Velocity Through Tubes, Feet per Second							
		0.5	1.0	1.5	2.0	3.0	4.0	5.0	6.0
Stanton	$\frac{1}{2}$ -in. vert. copper up	...	325	...	400	465	520	550	...
Stanton	$\frac{1}{2}$ -in. vert. copper down	...	420	...	470	525	560	585	...
Nichols	$\frac{3}{4}$ -in. vert. brass down	...	340	...	370	405	435	460	470
Nichols	$\frac{3}{4}$ -in. hor. brass	...	500	...	530	560	585	615	650
Hepburn	$1\frac{1}{4}$ -in. copper hor.	250	365	...	590	...	...	...	...
Hepburn	$1\frac{1}{4}$ -in. copper corrug.	360	560	...	...	...	...	...	...
Richter	$1\frac{1}{2}$ -in. hor. corrug.	460	...	...	...	...	...	...	...
Weighton ...	$\frac{5}{8}$ -in. plain tubes	...	380	...	615	760	865	940	...
Allen	$\frac{5}{8}$ -in. hor. tubes	...	225	...	290	365	...	...	...
S. & K.	$\frac{5}{8}$ -in. vert. brass U. & D.	...	400	...	590	710	810	880	945
Webre	$2\frac{1}{2}$ -in. vert. iron	80	153	212	260	...	...	...	...

In addition, on the chart (Fig. 1) are platted the following points as reported by E. W. Kerr.

K-1	$1\frac{3}{4}$ -in. brass horizontal	Velocity, 4.7	Transmission, 71.8
K-2	$1\frac{3}{4}$ -in. brass horizontal	Velocity, 4.7	Transmission, 137.5
K-3	$1\frac{3}{4}$ -in. brass horizontal	Velocity, 5.5	Transmission, 83.99
K-4	$1\frac{3}{4}$ -in. brass horizontal	Velocity, 5.2	Transmission, 375.1
K-9	$1\frac{3}{4}$ -in. brass horizontal	Velocity, 4.7	Transmission, 202.6
K-13	2-in. copper horizontal	Velocity, 2.0	Transmission, 69.67

Numerous tests by independent investigators have shown that, other things being equal, the transmission of heat is very much affected by the velocity at which the liquor travels through the heating tubes. As a matter of interest and study, the accompanying table and chart have been compiled from published data taken from various sources. The outstanding feature of these data is the apparent total lack of agreement among the authorities on the subject probably due to neglect of proper consideration of the basic elements brought out above. To the unbiased student, it seems that two conclusions are justified:

1. Generally speaking, increased velocity of circulation is accompanied by increased heat transmissions, but the shape of the curves is subject to much variation, and a hesitancy is felt in using an expression which would describe these functions;

2. The evidence points to higher transmissions as the tube diameters are reduced, although there are certain conflicting data.

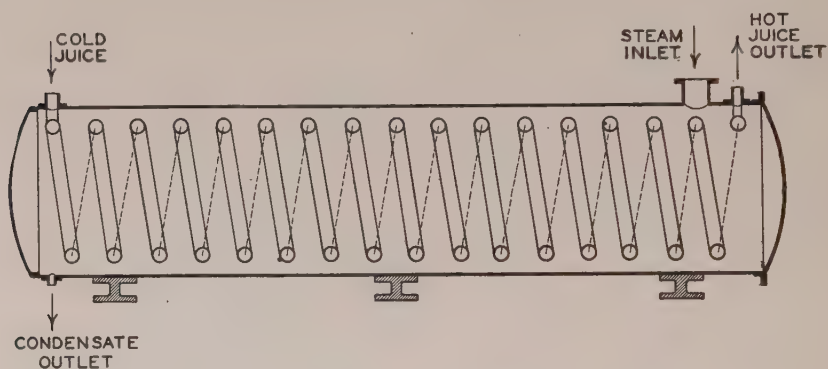


FIG. 2.—Old Type Single Coil Cylindrical Liquor Heater.

So far as the designer is concerned, in view of the facts presented, he must be satisfied to adhere to recognized broad principles outlined, to be guided by actual field results. The points established in the chart (Fig. 1) indicated as the K series are actual observations under actual operating conditions with commercial apparatus in the cane sugar industry, as reported by E. W. Kerr, who has already been quoted, and they show how dangerous it is to trust experimental data.

The viscosity of the liquid being heated has a marked bearing on the work of heaters, and is discussed more fully elsewhere. (Chapter 3.)

There are in use a number of types of heaters which it is well to consider in detail. The illustration (Fig. 2) shows what is commonly called the coil heater. Here, the liquid being heated is pumped through a very long cylindrical coil enclosed in a container into which steam is supplied. The liquor velocity is very high, as it has been found that by this method, it is possible to avoid completely the scaling of the surfaces on the liquor

side, which is scoured clear by skin friction. In fact, this is carried to such an extreme, that the coils actually erode after a comparatively short time, their life being about three or four years. From the nature of the case, successful operation of this heater would be impossible without such high velocity where very scaly solutions are heated, for the surfaces are inaccessible for mechanical cleaning to which it is often necessary to resort, where the obstinacy of the incrustations is such that they will not yield to chemical treatment.

The head required to force the solution through this heater is very great, sometimes as high as 100 to 125 lbs. per square inch, and consequently, the power expense is considerable. The heat transmissions are relatively low as explained before on account of the inherently bad conditions of large tube diameters and long lengths, both of which increase the liquid film, and consequently decrease the overall coefficient.

This design has until recently been very extensively used in Cuba. As can be seen, no trouble is experienced from expansion and contraction on account of the construction which provides easily for all such changes without undue stress.

In Fig. 3 is shown a vertical one pass heater which has been used a great deal in the chemical industries.

The steam side of this heater is very efficient, having good circulation through a continuous path of reducing cross section, such as has been recommended in the discussion above. This path is obtained by the installation of several baffles one above the other on opposite sides of the shell, the reduction in the section of the path being obtained by properly spacing the baffles. The baffle plates have a slight down pitch towards the other periphery, and thus catch the condensation as it falls from the surface above, throwing it against the shell of the heater, where it drains to the bottom without flooding the tubes and increasing thereby the steam film resistance. The non-condensable gas and air are removed from the bottom under the last baffle, where the incoming liquor is coldest, thus reducing their volume to a minimum. The condensate is also removed from the bottom at the point shown. Sometimes, where the volume of steam is great, due to sub-atmospheric operation, there is an annular space at the steam inlet, in order to provide ample entrance space for steam, otherwise, it would be necessary to leave off some of the tubes immediately in front of the inlet, which would have its cross section blocked by them.

The liquor side of this heater is as inefficient as the steam side is efficient. The cold liquid enters at the bottom, and goes up through the tubes in parallel, and thus the velocity of travel is very sluggish indeed, and with large diameter tubes, is ridiculously slow, so that poor heat transmissions are inevitable. The only feature to recommend this construction is the fact that very little head is required to force the liquor through, perhaps only a few inches. It is of course very accessible for mechanical cleaning, of which it will have much need, unless the operating conditions are unusually favorable. Ordinarily, short tubes are provided in order to avoid the necessity of an expansion joint in the shell.

The vertical heater shown in Fig. 4 represents fairly good practice. The steam side shows no baffle plates, which have been omitted in early practice, but should be provided along the same lines as fully discussed in Fig. 3. It would also be better to provide separate outlets for venting and condensate as in the former case. They were left out of the illustration for the reason that it was desired to show the construction as it is usually encountered in practice.

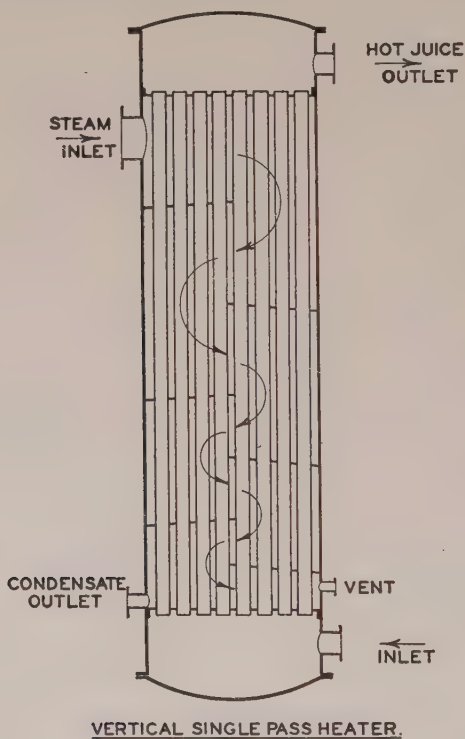


FIG. 3.

This particular unit has 96 tubes, and the liquid is compelled to pass through 6 tubes at a time from top to bottom, and bottom to top alternately, by having cells or compartments at the upper and lower heads which enclose groups of tubes so as to accomplish this purpose. There are hinged covers which are opened for cleaning, inspection, and repairs as the occasion demands. With short tubes, not over 11 ft. long, no provision need be made for expansion and contraction, as the deflection of the tubes provides the deficiency. This has been established by long practice. With greater lengths, it is necessary to take care of this feature, and this is usually done in one of three ways:

- 1) By providing stuffing boxes or ferrules at one or both ends of the tubes;
- 2) By putting an expansion joint in the body of the shell;
- 3) By enclosing one of the heads with its cover in the shell, permitting free longitudinal movement, or so-called "floating head."

By the last method, the accessibility of the equipment is impaired, and

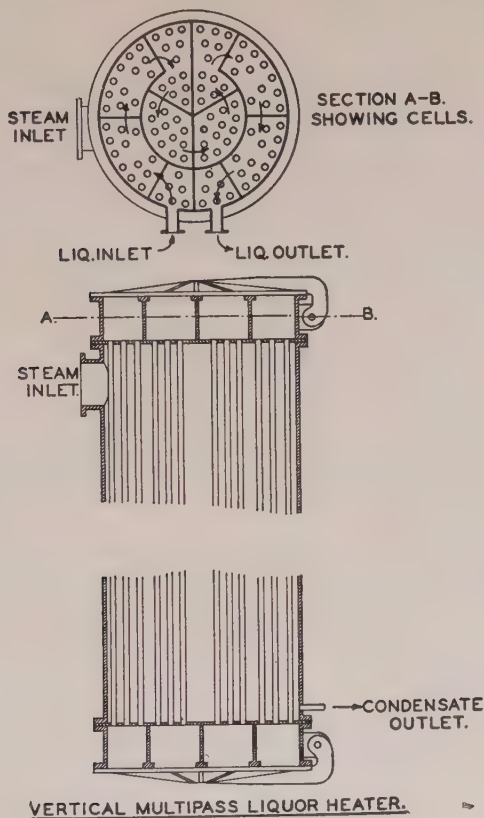


FIG. 4.

in order to make an inspection of the floating head, two joints must be broken, one in the shell, and one in the head. In case of a leak in the joint between the floating head and its cover, this will not be readily detected as would be in the case of the other types of construction.

The vertical multipass heater has both advantages and disadvantages from the operator's standpoint. So far as ease of cleaning is concerned, it leaves little to be desired, for both doors being opened, the tubes can be cleaned from the top, the dirt, scale, and wash water falling into a drain

below. Thus there is no untidiness on the operating floor, and the tubes can be washed out with a hose easily. The other valuable feature is the small amount of floor space required, the ease with which the piping can be arranged, and the general accessibility from all standpoints.

Especially when provided with rain plates, as in Fig. 3, the flow of liquid and steam are not counter current, and so, it will not be possible properly to cool the air before removal, and the volume will therefore be greater.

With a heater such as shown in Fig. 3 when in operation, because of the up and down flow, unless proper provision is made for venting the cells at the top head, difficulty is sometimes experienced in expelling the entrapped air, which becomes sealed at the bottom of the down tubes, especially if the rate of circulation is relatively slow. This results in increasing the head necessary to pump through the heater.

Another defect is in the location of the cells. The heating being progressive as shown by the arrows, the tubes in the inlet cell will be relatively cold, whereas those in the outlet cell which is adjacent will be at the maximum temperature, thus creating a maximum condition of unequal expansion, with the danger of troublesome leakages, unless stuffing boxes are provided to take care of these changes.

The same defect will exist if the heater is operated in a horizontal position as is often the case.

In Fig. 5 is shown the author's idea of a good simple design. Here, the heater is operated in a horizontal position, and rain plates are provided which serve the double purpose of removing the condensate drip, and providing a good path for steam with ultimate drain and vent at the bottom at the end of the path.

The heads are very simple with cells of straight line construction, the heating is progressive from the bottom to the top, the flow being by rows, forwards and backwards from the front to the back head, and thus adjacent tubes have only a slight difference of temperature.

The flow of liquid and steam are counter current which gives the best results from all standpoints.

If the tubes are not over 10 to 11 feet long, no allowance for expansion need be provided unless the ranges of temperature are abnormal. For ordinary conditions of heating from room temperature to ebullition or slightly above, no trouble from expansion will be experienced.

In multi-pass heaters, the arrangement of cells, size of tubes, and number must be determined from the particular problem in hand. It is customary to use fairly high velocities, from 5 to 6 feet per second for liquid speeds, under which conditions the fouling will not be serious.

Heaters must be cleaned often and installations must be in duplicates so as to permit this cleaning without shut down.

For effective commercial work, it is not safe to figure on transmissions much over 200 B.t.u.'s per square foot per hour per degree difference of temperature, and, unless the units are well designed and kept clean, this figure will dwindle down to less than 100. These transmissions are considerably lower than those obtained in surface condensers, but, on the

other hand, surface condensers generally use much smaller tubes, which, as was pointed out, seem to show better heat transfers.

In estimating the mean temperature of the liquid in its path through the heater the logarithmic mean, and not the arithmetic average should be used. The error of the arithmetic mean is not great if the range of temperature is low.

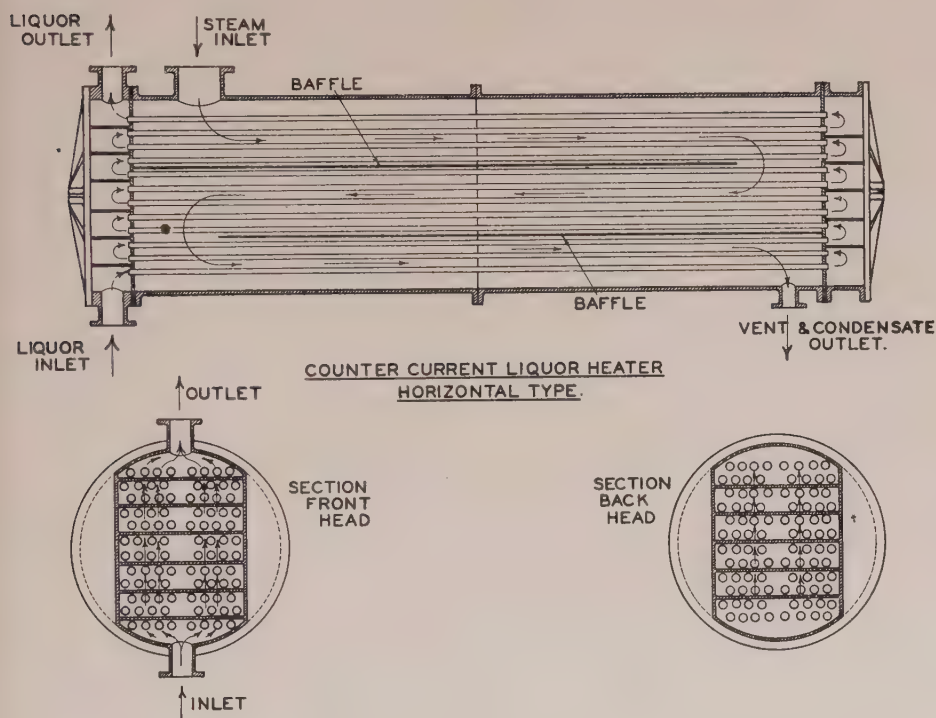


FIG. 5.

It is well to bear in mind that if steam used for heating is at a pressure below the atmosphere, more liberal heating surfaces should be provided, as air binding will come into play making high heat transmissions far more difficult, although there will be less trouble from fouling as a rule.

As a final precaution, it is well to emphasize the rigid necessity of keeping the surfaces of heaters clean for efficient results, and the period of such cleaning operations can only be determined in the field for each individual case, according to conditions encountered.

SECTION II.

Chapter I.

Information on the Operation of Evaporators.

In the preceding section, the general theoretical considerations have been discussed, mainly from the standpoint of the function of evaporators, their design, and their selection for a given problem. It is now in order to look at them from the practical operating standpoint.

The so-called "thermal engineer," who has charge of the execution of the design and plans, will not be the one to run the equipment. He will perhaps pay a casual visit to see that everything is in good shape, but the man in charge at the factory will probably be an ordinary operator who, if he has had any previous experience, will arrive with a swollen chest, his mind loaded with "dope" from his previous job. Some of his information will be surprisingly good, having been worked out from hard knocks at the cold-blooded school of practice, where alibis are not entertained. A large part of his data is of purely local origin and application, and the first detail to be attended to in starting new equipment, is to see that this man is properly informed, and not permitted to flounder around in his own way, thus basing this work on misconceptions founded on ignorance, for once he establishes his routine, be that right or wrong, he will invariably revert to it when left to himself.

The best scheme is to establish a simple schedule of operation arranged to take care of the specific local conditions as they are found to exist, and which will therefore differ considerably in different plants. This information must be couched in simple words, and be easily within the grasp of the men who will run the apparatus.

The class of labor used in this work is, as a rule, not of the best, and they must not be expected to do much thinking for themselves, not being familiar with the principles involved. It therefore pays to be very careful with these men in the beginning, and such attention will be rewarded in the end by successful operation.

It is really easier to teach a green man to follow out proper operations than to start with one having had considerable experience without careful instruction, and this is doubly true if the apparatus at hand differs materially from the one with which he has had experience.

In the following pages, an outline will be given of the various considerations necessary in this work, amplifying them where necessary.

Chapter 2.

Charts for Evaporators.

After many years of experience, it has been found that the best method of studying the work of evaporators is by making a picture of their operation, which visualizes the changes in their work as it progresses from one clean out period to the next.

In making such "charts," it is best to use the time element as the abscissa, and temperatures and other variables as ordinates. The apparatus must be properly provided with thermometers indicating boiling liquor temperatures in all bodies, as well as reliable vacuum gauges and pressure gauges indicating the conditions at all important points, whose readings are converted into corresponding steam temperatures to be platted on the chart. Readings should be made every hour on all indicating instruments as nearly simultaneously as possible.

As a measure of the rate of operation, the condensate from the steam belt of the first effect should be accurately metered. The meter need not necessarily be of the recording type, but should indicate the rate at which steam is being condensed in pounds per hour. Where there are no such instruments, a very satisfactory substitute can be made up in the form of a triangular weir, in which the head is calibrated to read directly pounds per hour. This weir should have a fair stabilizing tank in front to avoid surges which generally take place in the flow of condensed steam, as otherwise, it will be impossible to determine a fair average.

With new equipment, operating under untried conditions, the determination of the permissible period of operation from washout to washout is one requiring care and judgment, and there is no better way to decide this critical question than by means of a chart, which must be watched very carefully as the operation progresses. The limiting factor is usually the rate at which scale and dirt accumulates on the heating surface, and the rate of steam consumption will be an almost direct indication of this condition, as such scale retards the transmission of heat, and thus slows down the operation progressively as the accumulation increases, until the capacity of the apparatus becomes so limited that it is no longer capable of accomplishing the work necessary.

It is best, as a rule, not to allow the operation to proceed until this point is reached, for, as will be brought out later, it is much easier to clean out two small deposits of scale than one twice the thickness. In this work, the length of time from clean out to clean out varies from 16 hours to 15 days, the character of the solution being evaporated, and the design of

the equipment being the determining factors, so that it is evident that each case must be studied on its own merits. Even in the same industry these conditions are subject to much variation.

The cleaning out operation should be done thoroughly, rather than leaving small amounts of residual dirt to be carried over into the next cycle, as this accumulation becomes worse and more obstinate at each lap, finally necessitating a long and expensive shut down in order to clean in such a manner as to bring the capacity back to normal.

The intense fouling may be in any, several, or all of the various bodies, according to conditions, and it is impossible to predict in advance just

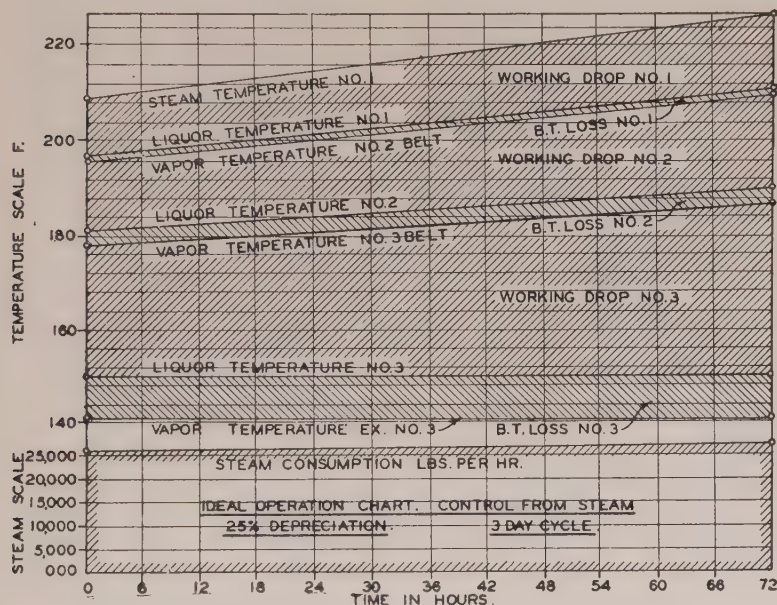


FIG. 1.

where it will take place, unless all the facts are known. For instance, in the sugar industry, on a quadruple, it may be in the first, third, or fourth bodies.

In order properly to illustrate, sample ideal charts are submitted based on the operation of the triple effect discussed in the chapter on Vapor Heating, Cycle 1. A three-day period has been arbitrarily selected, and a depreciation of 25 per cent. in the coefficient of heat transmission from the beginning to the end of the cycle. By this is meant, that the maximum capacity of the evaporator at the end of the period will be 25 per cent. less than it was at the beginning.

As to the method of controlling the work of the equipment, there are three methods, for each of which there is a chart plotted.

The first, shown in Fig. 1, is based upon maintaining the quantity of evaporation constant, and for the specific case mentioned, at 75,000 lbs. per hour. This is done by carrying the maximum vacuum at 24 inches, and throttling the steam supply so as to balance the heat to the load. The result will be that at the beginning of the three-day period, the pressure temperature in the steam chest of the first effect will be reduced by one quarter of the *net final working temperature difference*, or in this case, 18° . The contemplated maximum steam temperature being that which corresponds to 5 lbs. gauge pressure, or 227° , the steam temperature at the be-

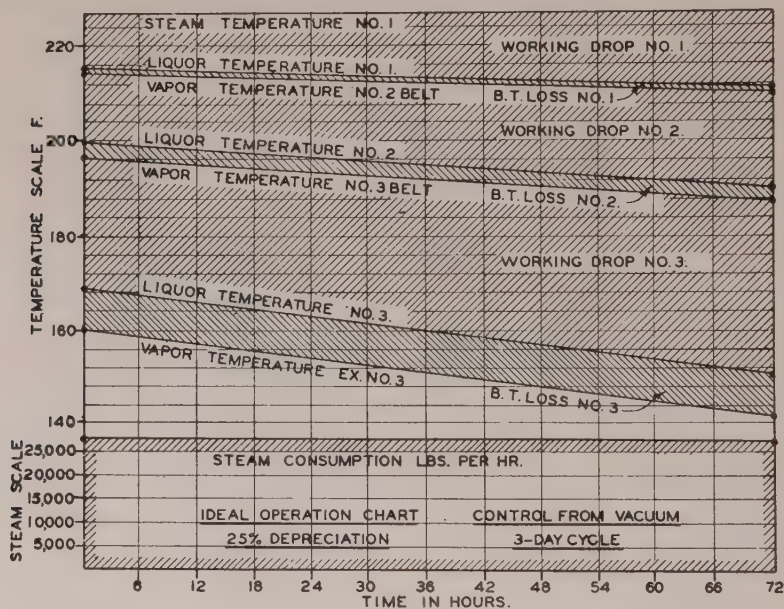


FIG. 2.

ginning of the operation will be $227 - 18 = 209$, as shown in the chart.

In order to maintain the rate of work prescribed, this temperature will be gradually increased as the operation proceeds, to compensate for the ever depreciating coefficient due to fouling, reaching its final and maximum, 227, at the end of the run.

Where there is a constant load, this is the best and most economical method, for when there are scale forming salts in solution, in view of the lower average temperatures, their precipitation and deposition on the tubes will not be so rapid. It is also true that under these operating conditions, there is a slight decrease in the steam consumption, as will be revealed by a heat balance. The actual steam consumption at the beginning of the period will be 25,900 lbs. per hour, as compared with 27,200 lbs. at the

end, a difference of 1300 lbs., and the average for the run will be 26,550 lbs.

The second method of control also presupposes a constant rate of operation as above, and will be the same for this specific case, 75,000 lbs. per hour. Fig. 2 shows the method of operation. Here, steam is maintained at a constant pressure and temperature at the point fixed, 5 lbs. gauge, and 227° , and the work is held down by lowering the vacuum in the last body, so as to reduce the net working temperature drop by 18° as before, thus bringing up the temperature of the vapors from the last body from 141° to 159° , corresponding to a drop of vacuum from 24 inches to about $20\frac{1}{2}$ inches referred to a 30-inch barometer.

When there is a shortage of water at the plant, this is a very advantageous method of operation, inasmuch as it materially reduces the requirements of the condenser, by permitting a wider range of heating, and therefore a smaller quantity of water.

There is also another advantage in that the condensation from the steam belt of the first effect is not under vacuum as in the preceding case, and the water leaves at 227° , a good temperature for boiling feeding.

The disadvantages are, in the first place, quicker fouling on account of higher prevailing temperatures, and poorer steam economy, which is shown by a heat balance to be 27,900 lbs. per hour. The final steam consumption being 27,200 lbs. per hour, the mean will be 27,550 lbs., or 1000 lbs. more than in the other case.

There is still another method of operation as shown in Fig. 3. Here, the full pressure and vacuum are maintained at all times, and the evaporator is allowed to run at its existing maximum capacity during the full period of operation. Where evaporators are short of capacity, this method is frequently used, and, as can readily be seen, a considerable increase is obtained. The equipment at the beginning of the period will evaporate as much as 100,000 lbs. per hour, the final being 75,000 lbs., and the average 87,500 lbs., or $16\frac{2}{3}$ per cent. more than in the other case.

The fouling will not be as rapid, as it has been well established that the faster an evaporator operates, the less the fouling, on account of increased speed. The economy presumably is constant, and when corrected for capacity, would give a steam consumption of 27,200 lbs. per hour for 75,000 lbs. of evaporation, a point between the other two averages.

Some operators accomplish the control of capacity by throttling the vents to the various steam belts, thus allowing them to become air bound. This is very bad practice, for if the non-condensable gases given out by the liquor are corrosive, as is often the case, there is grave danger of pitting the tubes.

These charts have been platted on the assumption that the depreciation of the coefficient in all bodies will be proportional, and, as stated before, this is not necessarily true. The shape of the curves would of course be modified to suit local individual conditions as encountered.

In actual work, the lines in the chart will never be straight lines as shown, but will oscillate up and down to correspond to the variations

which are always encountered. For instance, the working drop in any body will be increased by carrying too high or too low levels, or by fouling, or by air binding, or any other cause, such as irregular fouling. This is really the advantage of these charts in that they are very sensitive, and respond at once to the least irregularity, thus providing a means of locating the difficulties quickly and surely.

The information in these plats can be supplemented with explanatory notes, where necessary, and when there are important developments. It

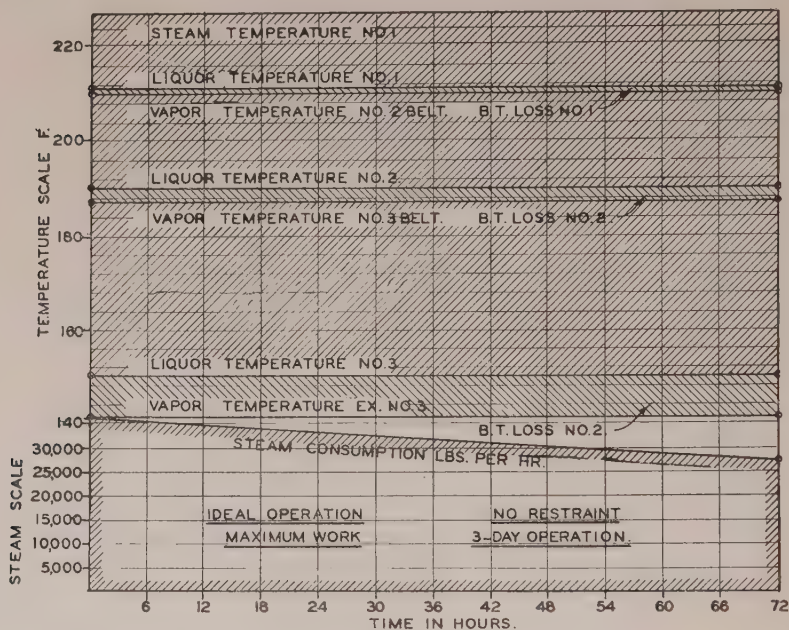


FIG. 3.

is also sometimes advisable to plat other data simultaneously, such as the operating levels, where hand control is used in feeding.

It is advisable not to use too small a scale, and 10 × 10 cross section paper, 20 inches or 24 inches wide is recommended. This can be secured in rolls and cut off to suit the length of the period of operation. The field of the chart should be cross hatched in different colors, which makes the data much more vivid and understandable.

Where very important operations are carried on, continuous keeping of charts filed for ready reference and comparison is recommended. When a good run is made, such a chart can be used as a standard by which to judge the operation at any time, and the information thus revealed is almost invaluable in case of difficulties.

Men trained to use charts soon learn to diagnose symptoms with remarkable accuracy, and are thereby enabled to direct the work properly with maximum efficiency.

When continuous charts are not justifiable, the work of the apparatus should at least be so recorded for a short period at the beginning, and the work of the evaporators checked occasionally, when there is need for it, or when unsatisfactory conditions arise.

Chapter 3.

Starting Up.

After completing the erection of an evaporating equipment, the next step is to apply a thorough water test to determine the tightness of all joints and piping.

Each steam compartment should be tested under water pressure by completely filling it, and subsequently putting on a pressure somewhat above that for which it was designed. All of the air must be released before doing this, so that the leaks will be indicated by water, and therefore easily located for correction afterwards. In vertical tube evaporators a leak in the lower tube sheet is much more serious than in the upper one, for the former will admit condensed steam into the liquor being evaporated, whereas the latter will merely allow a small amount of vapor from the steam belt to shunt through, a matter of small consequence, but of course to be avoided.

After having tested all the calandrias separately, the liquor side should then be filled as high as possible to locate and correct any external leaks, not only in the evaporators themselves, but in the entire system, which should be water tight.

It is also advisable to pass water through the various feed pipes, condensate pipes, drain pipes, vent pipes, etc., to make sure that they are all clear, for it very often happens that erectors, intentionally or not, leave material in them, such as waste, bolts, putty, and the like. We know of a number of cases where disgruntled men have deliberately resorted to this as a means of revenge. On a large project, there are always such men, and one must be on the lookout for little details of this sort, *taking nothing for granted*. Usually, the engineers who designed the apparatus can be given credit for knowing their business, and from experience, they proportion parts and pipes of ample size for the work, and therefore, if on starting operation, these seem to be insufficient in size, look for obstructions, as suggested above. We know of one instance where an entire automatic feed system was thrown out and replaced, only to find that the original pipes had been plugged with sacks.

After the water test, a vacuum test should be applied. It is best to try out one body at a time. To do this, close all valves, and raise vacuum on the last effect. If the apparatus is provided with a barometric condenser, be sure that there is enough water in the hot well to cover the end of the pipe, even after part of it has been drawn up in the barometric column by vacuum.

Unless there is a very large opening, vacuum should begin to rise as soon as the vacuum pump is started. Generally speaking, the leaks can be heard, and thus located, and chalked for correction. After locating a few leaks, do not stop the pump, but allow it to run, letting the vacuum take care of itself; in the meantime going over the entire equipment, and chalking up bad spots as they are detected. Where the vacuum losses come from improperly tightened joints, take up on them at once. On a dry apparatus, *i.e.*, one not in operation and not filled with liquid, a vacuum of 27 inches shows a fairly tight system, and anything much over shows very good conditions. Provided condensing water is available in sufficient quantities and at a low enough temperature, in actual operation, it is usually possible to get better vacuum than on the test.

After correcting the leaks located as above, if it is still impossible to get a satisfactory vacuum, the apparatus must be gone over with a candle. Wherever there are small leaks, the flame will enter due to the difference between the internal and external pressure, thus giving the exact location.

Sometimes, even after all apparent leaks have been taken up, it is still impossible to raise sufficient vacuum. This may be due to loose grained castings. An easy way to overcome this trouble is to paint the entire equipment with fairly heavy asphaltum paint while under vacuum, so that the open pores will fill up with paint, which soon hardens. Litharge and glycerine is excellent for small leaks, and a paint made in this way quickly hardens, making a tight coating.

After making the last body tight, vacuum is raised on the preceding, and the entire procedure gone over again. It will not be possible with a given vacuum pump displacement to raise as much vacuum on two bodies as on one, nor as much on three as on two, and so on. Also, it must be remembered that leaks are more harmful on the last body than on the preceding ones. It is, however, essential to have a practically tight apparatus. This is true of all bodies. If the pressure bodies are not tight, they will leak outwardly either liquor or vapor, according to the location of the hole, making an untidy mess, doubly objectionable if the solution being evaporated is of a corrosive nature, such as caustic soda.

When time is available, a run should be made on water before attempting to concentrate liquor, thus testing out the operation of the apparatus with all its accessories, including pumps, condenser, feed pipes, condensate pipes, etc. One generally encounters difficulties with the removal of the condensate from the calandrias, particularly on the last body. The concentrated liquor pump also tends to be troublesome. These two have similar obstacles to overcome, namely, poor operation on account of air leaks from the evaporator to the pump, either in the suction pipe, or in the pump stuffing boxes. This has been discussed somewhat in detail under the head of condensate removal. A little tinkering and use of common sense will generally rectify these difficulties, which are not of a complicated nature. They should amount practically to nothing but taking up leaks, if the pumps have been properly selected, and installed in accordance with suggestions.

It is well to remember that mechanics and erectors have one prevailing fault, carelessness, and one may expect to find upon starting operation a large quantity of foreign material, such as waste, nuts, bolts, matches, putty, tin cans, pieces of paper, and scraps of all kinds passing through the feed pipes and condensate pipes, washed down from the evaporator by the flow of liquor and condensate. All this has to be removed or it will interfere with the free flow through the piping systems. The strainers ahead of the pumps have to be cleansed a number of times before getting clear of these obstructions during the first few hours of operation. Sometimes, it may be necessary even to take down pipes which have become choked in this way. The main object of the water run is to ferret out all these little troubles, and rectify them beforehand.

Sometimes, it will be found that the "ammonia pipes," or air vents, dip down too deep into the steam belts, and so pick up water from the bottom tube sheet, in which case, they must be shortened. This can be easily determined by observing the discharge of the air header, or even by the sound of the flow in this pipe. By shutting and opening the individual vent pipes, there is a distinct difference in the sound if water is present, and thus the location can be readily determined, and corrected in time.

After the above preliminary work has been done, the evaporator is now ready to be put in full commission. There are certain variations in method of procedure according to type of equipment and material being concentrated. Assume in this case, however, a standard vertical tube evaporator operating on a non-crystallizing solution, and in order to follow out the general illustration, take the case of the sextuple effect for which the calculations were made in Chapter 8, Sec. I. Assume also that the apparatus is provided with hand control. The method of procedure used is as follows:

First, liquor is admitted into the various bodies which are filled to a point *slightly below the upper tube sheet*. The vacuum pump can be started at the same time, and the injection water turned on the condenser—not too much water to begin with as it is best to have a relatively small amount of water until the last body is boiling. The reason for this is very important, and will be discussed later. If the condenser is of the ejector type, however, it will be necessary to admit a sufficient quantity of water to fill the combining throat, otherwise the vacuum will not rise.

All the vent pipes on the various steam belts must be wide open to start with, and, if the evaporator is fitted with condensate pumps, these should be all started at once.

Steam is now turned on the calandria of the first effect, and as soon as the thermometer in the vent pipe indicates that all the air has been driven out, the control valve should be throttled down to about one-quarter turn as indicated in the discussion on non-condensable gas removal.

The feed valves in the piping connecting the various bodies should be all closed, otherwise, the liquor will all go towards the last body. Do not try to go too fast, or there will be difficulties.

In view of the fact that the liquor in the first effect is relatively cold,

there will probably be a vacuum in the calandria at the beginning, even if the steam valve is wide open.

In starting up a large evaporator of this sort, it is necessary to feel one's way. It is therefore unadvisable to give the full supply of steam at once, particularly in view of the fact that the proper settings of the valves are as yet undetermined, and that all the surfaces are clean, and therefore much more active than would be normally expected. The liquor temperature will now rise more or less rapidly, and the pressure on the steam side increasing at the same time. When this temperature reaches the point which corresponds to the vacuum or pressure on the liquor side, ebullition will begin, releasing vapor which must now push ahead of it the air contained in the vapor space of the first effect. All this air must be expelled through the vent pipe of the second steam belt, and in view of the fact that the size of this pipe is small, only 2 inches in this case, it will take some little time. A considerable part of the heat will be absorbed in heating up the metal parts above, the shell of the vapor belt, the dome, catchall, vapor pipe, and finally the calandria of the second body. In the meantime, inasmuch as the vapor is generated much faster than it is possible for it to drive out the air ahead of it through the small vent, the pressure here will tend to build up, or, if there is a vacuum, this vacuum will drop.

If there is a piston condensate pump on the steam belt of the second effect, part of the air will be removed by it, thus relieving the vent pipes of part of their excessive load.

If the equipment is arranged for by-passing the condensate, as suggested in the discussion on condensate removal, none of the air can pass through at this point on account of the water-sealed syphon interposed in the connection between adjacent bodies.

However, as soon as the zone of steam driving the air ahead of it reaches the heating surface of the second body, the tubes being filled with cold liquor, condensation will immediately begin to take place, increasing rapidly as the steam belt is relieved of air. This will at once relieve the pressure, or increase the vacuum, as the case may be, in the vapor belt of the first effect, and a flash will take place, lowering the temperature of the liquor at this point to correspond to the now reduced pressure or increased vacuum. The flash will be accompanied with a distinct swelling effect of the boiling mass, which will be more or less marked according to the surface tension of the liquid.

In other words, if the material is of a foamy nature, the swelling will be greater, as the bubbles so generated will be harder to break. It is to minimize the effect of this flash that the liquor level on starting should be below the top tube sheet.

The next step is a repetition of the above between the second and third effect. In the meantime, make sure that the water of condensation is being removed.

Adjust the valve in the vent pipes of the second steam belt to about one-quarter open, and allow the operation to proceed as before. It must be remembered that air displaced through the various vent pipes all goes to

the condenser, whence it is removed by the vacuum pump, whose capacity is naturally limited. One cannot expect, while this process is going on, the vacuum to rise very high in the last effect. This will not take place until the entire apparatus is in operation.

It is necessary to look after the level of the liquor in the first effect; for, as evaporation goes on, it will drop, and finally reach the point where circulation will no longer take place. Before this happens, the feed valve must be opened slightly, just enough to maintain the proper level.

As long as froth is visible above the tube sheet, there is enough liquor, and maximum operating efficiency is secured when the levels are maintained at this point and no higher. Any material increase of level is accompanied with a distinct decrease in the capacity of the apparatus.

Until the routine of operation and the setting of the valves are established for normal work, the control of the feed pipes is an operation requiring great care and attention.

Each successive body of the multiple will start in turn, and attention to the proper levels will now become a vital consideration, for, as the feed will be required in the latter bodies, the entire series of valves ahead will require new settings. Each effect should be provided with a gauge glass indicating the levels of the boiling liquid inside, and it is well, especially on starting, to tie a string on each glass, which for convenience and ready comparison, can be slipped up or down, showing the proper levels to be maintained as they are determined.

It is also well that the control valves in the feed pipes should have some indicating device showing the settings. After routine is established, this will be almost unnecessary, but for starting up, it is a great convenience. There are so many things to be watched at the same time, that convenience and simplicity cannot be carried too far.

One has to be very careful when the last body begins to boil, for, as soon as the cushion of air driven ahead of the vapor is completely removed by the vacuum pump and this vapor enters the condenser, there will be an instantaneous drop of pressure or rise of vacuum, the extent of which will depend on the amount of water admitted in the condenser, and the temperature of the liquid in the last body when this takes place. Naturally, being at the point of highest vacuum, and the lowest boiling temperature, *the volume of vapor released by the inevitable flash will be relatively large*, and this must be carefully guarded against, or it will be accompanied with a large volume of heavy froth, filling the entire vapor space, catchall, and vapor pipe, with attendant serious loss in the condenser. For this reason, at the critical point, the injection to the condenser must be kept down to a minimum. If the evaporator is provided with baffles in the vapor space, it is easily possible for this flash to wreck them, necessitating a shut down for repairs.

The best provision against this difficulty has been found to be in relatively low liquor levels in the last body before ebullition and full operation is established. Thus, only the small amount of liquor in the tubes will be heated, resulting in a relatively small flash. Another wise precaution is not to admit any heated liquor into this body from the preceding one until

it has started boiling. Neglect of this precaution will multiply the flash many times. If the level is above the tube sheet, the entire mass will be heated before ebullition due to convection currents, and here again the flash will become much worse.

After all bodies are in operation, the levels are carefully maintained, boiling "dead end," or without withdrawing anything from the last effect. Samples are withdrawn for weighing with a liquor tester, and the gradual change in gravity noted. When the desired point is reached, the concentrated liquor pump is started, and the level in the last body regulated by the operation of this pump. Perhaps the best method of controlling its displacement is by throttling its discharge.

If the density rises beyond the desired point, the rate of feed must be increased, and if it is too low, the feed must be cut down.

It is usually necessary to take the raw liquor at a given rate, as fast as it comes from some previous process. This fixes the amount of liquor being fed, and the density is then controlled by changing the amount of steam admitted to the first steam belt.

Sometimes it is advisable to maintain a constant steam pressure in the equipment. The speed of operation can then be controlled by raising or lowering the vacuum in the condenser. From the standpoint of heat economy, it is better to operate with the maximum vacuum, allowing the pressure in the first effect to vary accordingly, as suggested above.

There is considerable misinformation as to the proper distribution of pressures and vacua in the various bodies of a multiple evaporator. This is really a very simple proposition. Good work under any given set of conditions means that the evaporator is showing a high coefficient of heat transmission. This in turn merely means that the surface is transmitting the heat required with the smallest temperature difference possible. In other words, the smaller the temperature drop between the two sides of the heating surface, the greater the efficiency of that surface, which is the same thing as saying that the difference of pressure should be the minimum possible for the work being done. The proper vacuum in each body is, therefore, the maximum possible under the existing conditions, and the proper pressure, where it exists, is the lowest pressure possible, unless predetermined by other conditions.

With levels properly set and the evaporator under normal running conditions, the valves on the vent pipes should be opened to such a point that further opening does not increase the vacuum, or lower the pressure in the preceding body. It is to be remembered that none of these changes take place quickly, and sufficient time must be allowed in all cases for equilibrium to take place. It is also indispensable to make sure that no other condition is altered while the observation is being made, such as a change of level, a change of concentration, a change in the vacuum at the condenser, or a change in the quantity of steam admitted to the first effect.

The particular drop of temperature, or pressure, in any one body operating under ideal conditions will depend upon the work it is doing. If the rate of evaporation is increased, the drop of temperature must be

increased also, and conversely. Aside from this, many other causes will bring about corresponding pressure and temperature changes. For instance, progressive fouling of the heating surfaces will increase the temperature drop in proportion to the thickness of the scale. So will high operating levels, so will excessive densities, and many other factors.

For this reason, it is very difficult, except from long experience, to predict with accuracy the probable operating pressures and vacua for any given apparatus, for these will vary considerably even between cleaning periods, as the distribution of scale is never uniform in all bodies, and one must of necessity, be satisfied with a fair approximation representing what judgement indicates will be a probable fair average.

For operating purposes, however, adherence to the suggestions given above will give good results. There is no such thing as too much vacuum in the steam belt of any effect, or for that matter, in the vapor belts, except insofar as these vacuum conditions may bring about entrainment losses, where insufficient provisions have been made to guard against them.

It is very bad practice to lower the vacuum, or increase the pressure in any steam belt by closing down the vent, for in this way, the surfaces become gas bound, giving more intense boiling in some parts than others. In addition, such gases are of a corrosive nature, and by permitting their accumulation in this way, they have a better opportunity of attacking the metal of the tubes. Therefore, at all times, unless there is a very special reason to the contrary, the maximum vacuum possible at all points should be maintained.

The flash effect described in this discussion will vary with the design of the apparatus. The greater the liquor contents of any given evaporator, the worse the tendency in this direction. Certain evaporators having very small liquor contents, such as the long tube types, are practically unaffected.

In ordinary parallel current operation, *i.e.*, where the feed follows the direction of the bodies, flowing by difference of pressure from the first to the second to the third, etc., it is generally true that the bodies nearer the condenser have the largest temperature fall, mainly on account of increased density and viscosity due to the higher concentration and lower temperature. Sometimes, however, the drop on the first effect is greater than that on the second. This, usually, is because on account of the temperature of the entering liquor, this body is called upon not only to produce the vapor condensed in the second effect, but it must also heat the feed for the entire apparatus to its own boiling temperature. Thus, in the case of the sextuple effect in question, the first steam belt must transmit 14,320,000 B.t.u.'s per hour, whereas the second body only transmits 9,520,000. It is therefore logical to expect a larger temperature fall in the first body. If, however, the temperature of the feed entering the first effect is at or above its own boiling point, the temperature fall in this case will be less than for the second body.

When the evaporator is operated "counter current," however, in which the feed enters the last body, and is pumped back from this to the preceding and so on, leaving the first effect at its ultimate concentration, the temperature distribution will be quite different. This is a special method of

operation which is discussed elsewhere. It is seldom used except for special reasons.

Where level controls are provided, the problems encountered are very much the same, except that once properly set, the levels require no adjustment or attention, which makes efficient work certain and easy, and eliminates the danger from careless operators, particularly as regards the carrying of too high levels, a fault very common and very hard to correct, for the tendency of all operators is to carry high levels, and avoid thereby the necessity of careful adjustments necessary for efficient work. Neglect in this direction not only slows down the machine, but increases the danger of entrainment.

It must be carefully impressed upon the men in charge that the apparatus is placed in the plant to evaporate, and not to serve as a storage tank.

Chapter 4.

Shutting Down and Cleaning.

Practically all solutions deposit dirt or scale on the heating surface of evaporators, which, depending on the nature of the case, slow up the operation more or less, eventually necessitating a shut down for cleaning, the capacity having been so reduced that it is no longer possible to take care of the full load.

Depending upon the composition of the solution being evaporated, and the conditions of operation, the accumulation will be more in some bodies than others. This is practically always so, and in many cases, not only will the quantity be different, but the composition of the scale as well.

It is interesting to note the variation of intermediate pressures and vacua as the apparatus becomes foul. Operators soon become accustomed to these observations, and can quickly tell the condition of their equipment.

If the machine is clean, it generally has considerable reserve capacity, which is held down by running on a smaller temperature fall, either by reducing the steam pressure in the first body, or by cutting down the vacuum in the condenser, or both, if desired. As the apparatus fouls, it will require a larger and larger temperature difference to do the work, so it will be necessary either to raise the vacuum in the last body, or increase the operating steam pressure in the steam belt of the first effect, or both, until the reserve has been all taken up, when there will be no alternative but operation at reduced capacity, or a clean-out.

Good judgment and experience will save considerable trouble from this source. For instance, it is sometimes a very difficult job to remove a thick layer of obstinate scale, and requires time, expense and labor. In general it is bad practice to permit too great accumulation. Time and overall capacities are both gained by cleaning at shorter intervals, for it is much easier to remove two thin layers of scale than one twice as thick. For that reason, it is preferable where there is a marked tendency to accumulate incrustations to clean as often as practicable.

When ready for a shut down, there should be a regular method of procedure, or routine, which can be established according to plant conditions. Generally, it would probably be as follows:

1. Shut off steam;
2. Shut off feed;
3. Break Vacuum;
4. Liquidate;
5. Wash out with water;
6. Remove scale by approved method.

Some of the above steps require explanation. There are no comments on (1) and (2). It is not always customary to break the vacuum at once. If vacuum is not broken, all bodies will begin to flash due to the drop of pressure, or increase of vacuum on account of the smaller amount of heat to be transmitted in the condenser, as well as in all the preceding bodies. Thus, if under operating conditions, there be a pressure in the first effect, this will shortly turn into a vacuum, which will gradually increase as the temperature falls.

By taking advantage of this, it is possible quickly to liquidate into the last effect, from which the residual liquor can be removed by the concentrated liquor pump. Here again, we must caution the beginner not to go too fast, but to run this liquor to the last body only as fast as the pump will remove it, keeping the levels in all the bodies from rising too high, as, if this precaution is neglected, there will be considerable flashing with the consequent danger of "boiling over."

As this process of pressure reduction is going on, as soon as there is a vacuum in the first body, break it by opening to the atmosphere, and when it is completely liquidated, close the feed valve between the first and second body, and break the vacuum on the second; the same procedure is followed on the next step, and so on until the entire apparatus has been emptied.

After having washed out, the manheads should be removed for observing the condition of the tubes, after which the cleaning operation can be started. There are a number of distinct methods, the one to select depending upon the nature of the material to be removed.

(A) If the incrustations consist of soluble salts, or loosely attached dirt, it is merely a question of washing out thoroughly with water. It is best thoroughly to agitate this water by means of a perforated steam pipe properly located under the bottom tube sheet. This is really better than actually boiling the water, inasmuch as the mechanical agitation in this way is practically just as intense at the lower part of the tubes as at the top. By boiling, ebullition takes place probably only near the upper parts of the tubes, the levels being above the top tube sheet, and the amount of steam being relatively small. It would be best to do this operation under maximum vacuum, as, on account of the greater specific volume of vapor under these conditions, a given amount of steam admitted below the tube sheets will produce a very much larger volume, and therefore much better mechanical agitation at the same expense. After a given time determined by experimentation, the dirty water is removed, the manheads opened, and the tubes flushed with a hose to wash down any loose dirt which may be left.

(B) Generally, however, it is not such an easy matter, for the scale is usually of an insoluble nature. It is a fact that distilled water dissolves practically any material, and it is a matter only of relative solubility. If, therefore, there is distilled water in sufficient quantity, it is sometimes possible to remove by its use the most obstinate scale such as calcium sulphate, for instance. To do this there must be a considerable volume of water, which must be changed continuously while boiling out. Direct

steam agitation as suggested above is recommended. It has been possible by this method to remove a layer of "Gyp" $\frac{1}{16}$ inch thick in six hours' time on a fairly large evaporator. There is considerable distilled water available in the operation of all evaporators, for all the condensate, or sweet water, is distilled water, and it is therefore only a matter of providing storage. The success of this method will depend on whether or not the scale is occluded by gummy, organic matter, and, even when this is the case, surprisingly good results are obtainable.

(C) The usual method of chemical scale removal, where such is necessary, is to boil out the evaporators with a dilute solution of caustic or soda ash, or both mixed. Sometimes, this is followed by a treatment with muriatic acid. The treatment requires about ten hours to be properly done, depending upon the obstinacy of the scale. Caustic seems to attack organic matter and soften the scale, whereas the acid has a solvent effect. Usually, after this kind of washout, there is loose scale found in the bottom of the evaporator, some of it dropping down even after starting up with liquor. It is necessary, therefore, to make provisions against choking the feed system with this accumulation.

(D) Sometimes hard scale is so unyielding, that chemical treatment is of no avail, and nothing remains but mechanical scraping or boring. This method should be used only as a last resort, for invariably, the material of the tubes is scratched forming anchor spots for further and more rapid accumulation of scale when the evaporator is again placed in commission. The wear and tear may be considerable, and very often the tubes are damaged.

(E) Some evaporators, such as small distilled water apparatus, have their surfaces so arranged that pressure variations cause a change in the shape of the tubes, and this, of its own accord, cracks off the scale. This is a very desirable feature, and will probably be used in large equipments in the future, where conditions permit.

(F) The Lillie Evaporators use a method of reversing the order of the bodies, both as to vapor and as to liquor, which it is claimed, eliminates the necessity of cleaning under certain conditions. There are undoubtedly certain merits in this contention, but it is probably not infallible. There are also complications involved, such as the use of two condensers, large vapor valves, and the pipe mechanism.

Chapter 5.

Scaling and Fouling.

The most serious problem we have to face in evaporation work is the depreciation of the coefficient of heat transmission due to the fouling of the heating surfaces, either by the accumulation of scale or extraneous matter in practically all solutions to be evaporated, which is sometimes further aggravated in the case of crystallizing solutions by the formation of salts which adhere to the tubes, and build up, especially if the apparatus is not provided with mechanical circulation.

It is fairly easy to estimate what an evaporator will do when it is clean, but it is an open question always as to how long it will carry on even at a much reduced rate when the contamination begins to take place.

It will be illuminating to refer to work which has been done on the subject. Professor E. W. Kerr, formerly at the Louisiana State University, about ten years ago conducted a series of experiments which brought out some of the most valuable actual data we have to-day, and his work has been generally quoted all over the world. These tests reported in *Louisiana Bull.* No. 149 were made in two series, the first being on small scale apparatus at the university, and later on, a large number of full-sized equipments were tested in actual practice under factory conditions.

The most important fact brought out in this work was that the heat transmission actually shown in the small scale experiments was several times as great on the average as that in actual practice, with the same design of equipments in both cases, the difference in size of bodies between the laboratory work and the field work being about one hundred to one roughly speaking. It must be admitted, besides, that the conditions as to temperature drops were not identical, but if they had been, there is every reason to believe that the discrepancy between the two series of experiments would have been, if anything, greater.

The radical difference as stated above, however, was that the experiments were conducted on a small evaporator, with clean heating surface, on a solution made to order from pure sucrose dissolved in distilled water, whereas the big scale work was done on defecated cane juice carrying with it all the normal impurities and scale-forming elements such as are encountered in the daily operation of a commercial sugar factory, sometimes after a period of six days' operation.

According to Kerr's report, he made two tests on one particular equipment in a factory, the first being three days after cleaning, and the other

after six days' work. The coefficients reported for the various bodies were as follows:

Bodies	No. 1	No. 2	No. 3	No. 4	Overall
After $3\frac{1}{2}$ days	321.2	507.76	203.17	90.01	201.97
After 6 "	238.5	270.17	97.11	79.45	129.70

Assuming that the depreciation of coefficient followed a straight line, the probable average coefficient of the equipment perfectly clean for overall transmission would have been about 303. The matter appears to the author as being so important that he wishes to amplify with another example. He recently conducted a test on a small single tube experi-

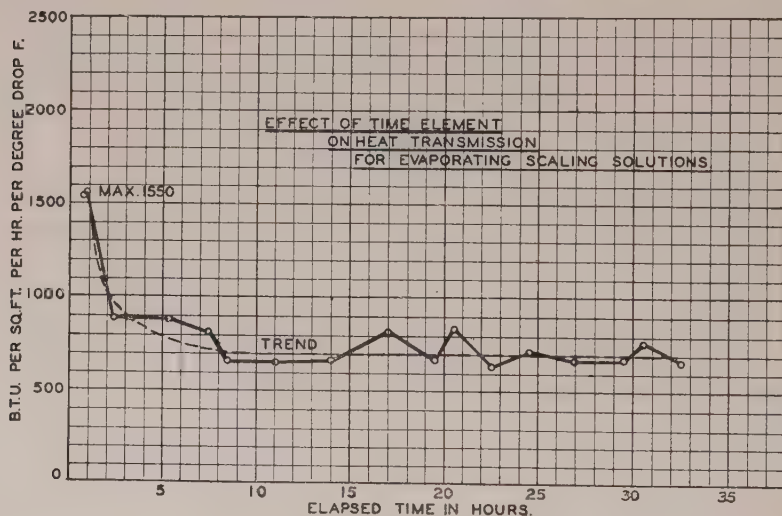


FIG. 1.

mental apparatus to determine the effect of scale on heating surfaces of distilled water equipments using sea water. The coefficient of heat transmission was determined every 15 minutes, starting with a copper bright tube. Fig. 1 shows the result, which to say the least may be surprising to one not familiar with this work. This test shows that at the end of two hours, the transmission of heat was cut in half, and that after this, the recession was progressive and continuous, but much slower. The experiment was repeated several times, so as to make sure of the data, and to the author's mind, there is no question of the facts. A careful examination of the tubes at the end of two hours' run showed a very thin scale, less than 1/100 inch thick.

These results are explained by the assumption that once a definite all-enclosing film of scale is formed, this coating is probably dry where it adheres to the tube, and the heat has to go through this layer of dry

scale permeated with slightly superheated steam before it can reach the water which it is to evaporate. It is unquestionably true to a greater or less extent that such conditions obtain in the average evaporator practically at all times.

There is no other single factor that will effect the work of evaporating equipment more than this, and it must therefore be rigidly taken into consideration in all cases.

Furthermore, any incrustations accumulating on the heating surfaces will act in very much the same manner, such as, for instance, the deposit of salt on the tubes of a brine evaporator, or the deposition of crystals from any solution on heating surfaces of such equipments.

In estimating the probable coefficient of heat transmission it is therefore necessary to make very liberal allowances for the fouling of the tubes, based on experience with the particular solution to be treated, or solutions of similar nature and behavior, and it is very dangerous to be guided by small scale experiments conducted under ideal conditions for small periods of time.

The counter argument would be that such surfaces must be kept clean. Quite true, but there is a well-defined limit in all industries as to the amount of time permissible for cleaning. In sugar factories, it is common practice to shut down eight to ten hours for cleaning once a week. In other industries, the practice has to be modified according to limiting conditions, and depending on the obstinacy of the incrustations, the rapidity with which they accumulate, and other considerations. In one case, the author found that it actually paid to clean out every day. The best method will depend on local conditions, the possible method of cleaning, and the cost of doing it.

Circulation has a marked effect upon the accumulation of scale. We have always found that the fouling was more serious at the lower part of the tubes than at the top. The velocity of travel at the bottoms of the tubes is of necessity sluggish, and the faster speed is at the top of the tube, the slower it is at the bottom. (Refer to Chapter 13, Sec. I, for more details.) The speed at the top of the tube is determined by the evaporative rate of the equipment, and the vacuum conditions in a given equipment. The matter of selection of equipments is therefore a vital consideration, related to the fouling tendencies of the material to be evaporated.

In many instances, the scale-forming constituents consist of lime salts which are less soluble hot than cold, and it then becomes possible to eliminate these salts to a great extent by a preliminary heat treatment with subsequent settling, where such a treatment will not injure the other valuable materials in solution.

In some cases, the solutions coming to the evaporator have a temperature considerably below that of the boiling point in the first effect. Where there is a marked tendency to foul, the problem is very much aggravated under such conditions, for the heating surfaces are called upon to do a heating operation for which they were not designed, and this heating generally takes place at the point where the circulation is most sluggish,

namely, at the bottom of the tubes. It is advisable to preheat to the boiling point before entering the evaporator, and where this is not feasible, we suggest spraying the feed in the vapor space, so as to accomplish this heating by direct contact with vapor rather than by the heating surface, thus the evaporator is called upon to make more vapors than normally which improves the circulation, makes the fouling less, and imposes on the evaporator a load for which it was designed, namely, evaporation.

Problems similar to the above have been discussed in the chapter on Shutting Down and Cleaning, to which the reader is referred.

Chapter 6.

Condensate Removal.

One of the most important details requiring special attention is the removal of water of condensation from the calandrias as fast as steam is used. It is a well-known fact that water is a very poor conductor of heat, and any accumulation simply blanks off that much heating surface, slowing up the apparatus. Indeed, it is very evident, especially with vertical tube apparatus, equipped with level controls, that as soon as the level of water begins to rise, the actual circulation, which is predicated on having the entire tube subject to the action of steam, will begin to diminish, and if the controls are properly set, when the condensate level rises to 4 or 5 inches, circulation will stop altogether, for the boiling mixture will not reach the top of the tube, and the liquor will simply boil away inside concentrating locally at a very much reduced rate, eventually stopping altogether, when the condensate level reaches the liquor level.

In addition, the "ammonia pipes," more properly called the vent pipes, which extend to within 4 or 5 inches of the lower tube sheet, will begin to draw out some of this water throwing it into the vapor space of the last body if connected to this point, or probably will become completely water bound if connected to the condenser.

It is therefore wise to make adequate provisions against the above-mentioned condition. In addition, there must be provided in each steam compartment a water gauge glass to indicate the level of condensate in it, and this glass must be so located and lighted that it is within easy and constant observation of the operator. The glass tube should have a white backing, so the water level indicated can be readily observed. Any deviation from the normal operating level in this glass should be investigated at once. In fact, after placing the evaporator in commission, it would be well to mark the tube by pasting a label thereon, indicating the normal water level.

As to the methods of removing the water, there is much to be said. If the steam compartment is under pressure, such as is generally the case with the first body of a multiple effect, the problem is very simple, merely requiring the use of a low pressure steam trap, a U-pipe, or some such device. Where the condensate is to be sent a long distance from the evaporator, a receiver pump with a float control is desirable. It is better for reliability of operation to put in an equalizing pipe from the top of the receiver to the upper part of the steam belt. This avoids the possibility of air binding.

Perhaps a better method is to remove this condensate with a centrifugal pump, and by making the proper connections, the operation will be just as satisfactory, whether the belt is under pressure or under vacuum. Fig. 1 indicates the proper connections. Until within the last eight or ten years, it was considered practically impossible to draw water from a vacuum by the use of a centrifugal pump, particularly if the temperature approached that of the steam from which this water was condensed. Now this is done every day. Indeed, the accepted method of withdrawing water from high vacuum condensers is by the use of a centrifugal pump located immediately under the condenser.

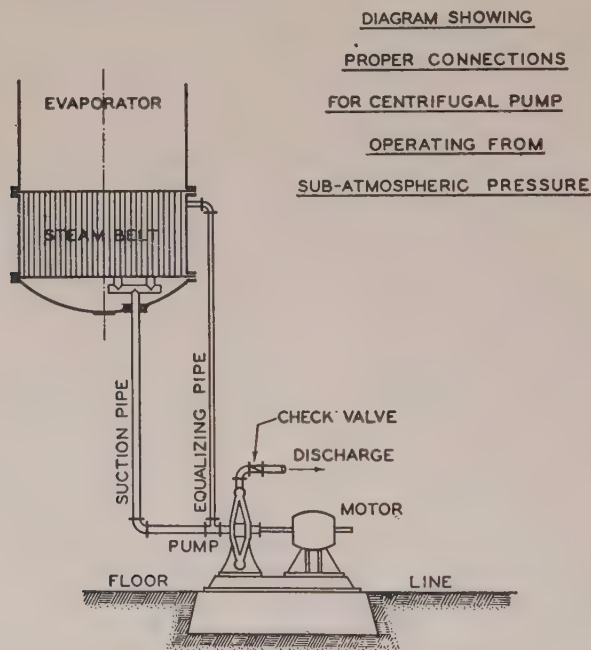


FIG. 1.

The essential point about the arrangement is to make the liquid enter the runner of the pump by gravity from which point it is pushed out by centrifugal force. A check valve must be provided on the discharge of the pump. The gravity flow is accomplished by use of a vacuum equalizing pipe which connects the upper part of the container from which the water is being removed to a point immediately above the entrance into the runner of the pump. This pipe must not be too small, not less than $\frac{3}{4}$ inch or 1 inch, depending upon the size of the pump. Its function is to withdraw air, vapor or gas from the centre of the runner, returning it to a point above, and permitting its replacement by liquid to be pumped. With such an arrangement, the operation is entirely reliable and satisfactory, although

it is wise to provide extra capacity, both as to head and capacity to take care of slower action caused by the presence of bubbles always found in the liquid removed under these conditions. It is also wise to select pumps equipped with lantern glands, or water sealed stuffing boxes.

It is important to emphasize the necessity of providing easily removable and cleanable strainers in the suction of the pump, if the liquid being handled is likely to contain scale or incrustations, such as is the case usually with the concentrated liquor pump on the last body. Without such provisions, the runners of the pumps will become choked, interfering with successful operation, and perhaps setting up off-set strains with possible injury to the pump.

In the old days, it was customary, when condensate was to be removed from steam belts under vacuum, to use for this purpose a wet vacuum pump, sometimes of the direct acting type, but more often of the fly wheel type. Edwards pumps are still popular for this work. One of the serious causes of trouble was the high temperatures of water and vapor which invariably gave trouble with rubber valves. This was overcome to a certain extent by a very foolish device called a sweet water condenser, which was merely a small jet condenser interposed in the suction of the pump, into which the hot water was cooled, presumably by diluting it with an injection of cold water. The result is apparent upon even superficial consideration. Not only was the "sweet water," or condensate, cooled, but also, a considerable quantity of steam from the belt above was condensed, resulting in just that much loss of heat, not alone in one body, but in all bodies so equipped. The use of the sweet water condenser was far from being universal, but in certain localities, it was considered indispensable.

In those days, it was not customary to have "ammonia pipes" and the function of the sweet water pump was to remove non-condensable gases as well as condensate. This of course was all right as long as these gases were heavier than steam, but if there was any quantity of relatively light gases, such as ammonia, which accumulated at the top of the steam belt, the sweet water pump became useless for this purpose.

The trouble with rubber valves was overcome by the use of brass valves instead, and this obviated the necessity of the "sweet water condenser."

The removal of condensate from steam belts of evaporators has always been a source of considerable trouble, for any small leak, which would not ordinarily be noticed, will introduce air bubbles into the suction of the pump. This not only causes the latter to become air bound, but in addition prevents the free passage of water downwards to the pump. It is therefore advisable to make such pipes as direct as possible, eliminating all unnecessary joints and turns. It is also wise to select ample sizes, so that relatively small leaks will not affect the operation.

Even with piston pumps at this station, there is much to be gained by installing equalizing pipes connected to the pump immediately below the suction valves. The piston pumps should also be equipped with basin heads providing a water seal for the piston rod stuffing boxes, thereby eliminating the possibility of air leakages. Pumps of the "suction valve-

less" type are much preferable. The well known Edwards pump is in this category. There is also a horizontal design, resembling a magma pump, which gives good results.

Where centrifugal pumps can be used, however, they are much preferable, being simpler, easier to keep up, and, if anything, more reliable. They run at constant speed, whether liquor is there to be removed or not, do not have to be governed, and, aside from occasional packing of stuffing boxes necessary to all pumps, require practically no attention. One of the appealing arguments to the man in the factory, who must be satisfied, is that they have no valves and no springs.

When the apparatus is set high enough, it is possible to remove the condensate by the use of barometric drains. In this case, the condensate pipes are simply led straight down, the ends being sealed in a trough or container below. Here, however, the net distance from the level of the water in the seal to the lower tube sheet above must be in excess of the hydrostatic head corresponding to the vacuum. Perfect vacuum corresponds to 30 inches of mercury, and this in turn corresponds to a column of water 34.5 feet high. It is therefore evident that for every inch of vacuum there must be a net drop corresponding to $\frac{34.5}{30} = 1.15$ feet, plus a margin over and above, which should be about 4 or 5 feet. Below is a table showing for different vacua the corresponding water head, and the advisable minimum difference of level between the lower tube sheet and the water seal:

Vacuum	Cor. Water Head	Minimum Drop Advisable
5-in.....	5.75 ft.	11 ft.
10-in.....	11.50 ft.	17 ft.
15-in.....	17.25 ft.	22 ft.
20-in.....	23.00 ft.	28 ft.
25-in.....	27.60 ft.	33 ft.

The use of barometric drains where the installation permits is extremely simple, quite satisfactory, and requires no attention. It is hardly necessary to call attention to the fact that the trough where the condensate pipes dip must be filled with water before beginning operation.

Perhaps it is well to call attention to one detail in connection with these gravity drains. In removing the water from the common trough, the direction of flow should by preference be from the last body towards the first. In addition, it is advisable to install dams or partitions making one compartment for each body, the height of these dams being such that the highest level of the water in the trough will be for the last effect, and the lowest at the other end, with successive drops for each compartment. In this way, the condensate will always flow from a cooler to a warmer section. The object of this arrangement is to prevent a flash back in the drain pipes, which might occur if the apparatus is stopped, and the vacuum broken, and started again, for under these conditions, when the water is drawn up in the condensate drain of the last effect, if its temperature is higher than that corresponding to the vacuum, a flash will start up, and

this water will actually boil back into the steam belt, filling it with water. Sometimes this problem has been off-set by dropping the seals into a deep well, making the column very much longer than provided in the schedule given above.

We have had occasion to witness this flash back, and the above suggestion was made to overcome it. There is no danger involved, as the water will eventually come to a balance with the vacuum, and normal draining take place. It is so easy to guard against, however, that we recommend the above suggestion.

There is really no good reason why condensate should be removed from each body of an evaporator separately either by the use of barometric seal or by the use of pumps. It is more economical from the standpoint of heat, to by-pass it from body to body removing the sensible heat at each drop by flashing as it passes successively into zones of lower pressures and temperatures, the final removal being done at the last steam belt. This results in a material heat economy, which is well worth while inasmuch as it really does not involve any serious complication.

This by-pass is accomplished by the use of loops or inverted syphons, whose length should be determined by the difference of vacua existing between adjacent bodies in accordance with the table given previously. Such a syphon should be provided with control valves on the discharge side as it enters the succeeding steam belt. In normal operation, these valves are left wide open. When starting, testing, or adjusting, however, they have to be manipulated.

Inasmuch as the quantity of water is cumulative, as it passes from one body to the next, the size of the syphon pipes must be increased progressively to take care of the larger quantity of water. The permissible velocity in these pipes is very low, the maximum being about 5 ft. per second, 4 ft. per second being good practice. This figure represents the actual water velocity at the inlet end. As the water approaches the outlet end, the pressure drops, a flash takes place, and the volume is greatly increased by the accompanying steam bubbles occluded in the mass.

Take a concrete example to illustrate. Thus, 1 gallon of water passing from the fifth to the sixth body of the sextuple, effect, whose heat balance is given in Chapter 8, Sec. I, on "Heat Balances," drops from a vacuum of 11 inches to 18 inches. This corresponds to a temperature fall of from 195.5° down to 169° F., and there is given up the difference between these two or 26.5°. The weight of a gallon being 8.33 lbs. the heat thus released is 221 B.t.u.'s. The latent heat of steam at 169° is 996.4 B.t.u.'s per pound, and the self-evaporation or "flash" is here $\frac{221.0}{996.4} = 0.222$ lbs. The specific volume of steam at 169° F. is 63.3 cubic feet per pound, and the volume of vapor so released from flashing down 1 gallon of water from 195.5° F. to 169° F. is therefore $63.3 \times 0.222 = 14.07$ cubic feet, or 105 gallons, which looks almost impossible, but is nevertheless true. The reaction, however, is not instantaneous, part of the flash taking place in the pipe, and part after leaving it. This discus-

sion shows why such pipes have to be made very large. Incidentally, the same conditions exist in the feed pipes of evaporators, where the flash on entering the last body is even greater in view of the greater temperature fall, and the much greater specific volume of steam under the higher vacuum conditions.

Perhaps the outstanding objection to by-passing the condensate is that this method introduces a much larger quantity of water into the steam compartments of the latter effects which may depreciate the transmission of heat. The answer is that with properly proportioned pipes, the amount of water actually in the steam belt at one time will be very little more. At all events, this method has been used for many years and does not show any reduction in the coefficient of heat transmission. Certain it is that a substantial economy has been accomplished as comparative heat balances will show, and the number of sweet water pumps has been cut down to one.

In several instances "flash pots" were used, or chambers connected with steam belts by a large vapor pipe, the arrangement being so disposed that only the vapor from the flash or self-evaporation entered the steam compartment, the water passing along and joining the condensate from this steam compartment before entering the next flash pot. The advantage gained, however, is so slight as compared with the results obtained by the other and much simpler method that the extra complication is of doubtful justification, unless the conditions make it indispensable, as in this particular instance.

It is well to note here that with the long tube evaporators, there is a distinct complication involved in by-passing the condensate, which has undoubtedly been the cause of serious trouble in the past, and still is in many cases to-day. It is about as follows: Evaporators of this general design are provided with vertical steam compartments relatively small in diameter, and very high. Steam is generally admitted near the top, and progresses in a generally downward direction, driving before it the remaining non-condensable gas, and the condensed steam. The water and non-condensable gas are by-passed from one body to the next through one pipe at the bottom, which in this case has no loop, but merely connects across from the bottom of one steam belt to the bottom of the next.

Here, as has been shown above, there is a considerable volume of vapor released, particularly in the last effect, by the flash at this point, which must be condensed by the heating surface of the evaporator. On the other hand, vapor is moving downward from above, and between these two sources of heat somewhere is a zone of vapor saturated with non-condensable gas, which to all intents and purposes, has no outlet. It eventually finds its way out by diffusion and eddy currents, but it is easily conceivable that this phenomenon causes considerable irregularities in the work, if not complete stoppage under certain conditions. The remedy is to apply the flash pot, allowing the vapor therefrom to enter the steam belt at a point very much higher up, thus not interfering with the free downward passage of the non-condensable gases. In addition, it seems that it would be preferable to remove the condensate in a separate pipe with

loops, and have an independent connection for withdrawal of gases at the bottom of the steam belt, at a point above the normal level of the condensate, thus avoiding the annoying interference of the other arrangement. In practice, such an arrangement gives entirely satisfactory results, and long tube evaporator so provided will operate with surprising smoothness and regularity.

Chapter 7.

Non-Condensable Gas Removal.

One of the earliest and most difficult problems encountered in the original work of designing and operating multiple effects under vacuum was the proper removal of air and non-condensable gases from the steam side of the tubes, and all of the early literature on the subject emphasizes it. In the early apparatus, it was sometimes customary occasionally to push the pressure in the calandrias up to above the atmosphere in order to be able to drive out these gases, afterwards re-establishing normal operating conditions as to vacua. The steam belts were provided with air cocks discharging to the atmosphere for this purpose. The surfaces so vented were of necessity always partially air bound, and for that reason could not be expected to be efficient.

To-day, with the knowledge at hand, attention to this detail is a matter of prime importance, and one worthy of care and study, just as it was then. Certainly, it is one of the first requisites for efficient heat transmission. It matters not if the gases come from air leaks in the vacuum bodies, or if they are gases liberated from the boiling liquid during the process of evaporation; for their presence even in small quantities is very detrimental to good work, in addition to the fact that the steam side of the tubes is very often corroded, especially when ammonia or carbon dioxide are present, for they are always accompanied by heat, air, and moisture, conditions known to promote chemical reactions of this kind.

The effect of the presence of non-condensable gases on heat transmission is two-fold. In the first place, if such gases exist in any material proportions, they reduce the temperature of steam or vapor by exerting a partial vapor pressure. It does not require a very large amount to lower the temperature of the mixture to such a point that condensation is materially reduced, or even stopped altogether in certain zones. These gases generally accumulate in certain spots, pockets or stratifications, and tend to cling to the surface of the tubes where they can do most harm, being driven there by the constant flow of steam in this direction. It is therefore quite a problem to get rid of them, and the real solution is to so design the equipment that neither zones, pockets, nor stratifications can exist. Nevertheless, there are many evaporators in use to-day not so designed, and the problem is therefore serious, and must be so recognized.

This can best be illustrated by an example. Referring again to the heat balance for the sextuple effect discussed in another chapter, consider the conditions existing in the steam belt of the last body. The temperature

of the liquid on the inside of the tubes is 134° F. The mean hydrostatic head increases this by 9° , making the mean temperature from the top to the bottom of the tube 143° F. If therefore, the temperature of the mixture of steam and gas at any point of the equipment comes down to 143° , ebullition will cease altogether. There was assumed a vacuum of 18 inches in this steam belt, corresponding to an absolute pressure of 12 inches. The pressure relations would therefore be as follows:

Vapor pressure corresponding to 143° F.....	6.34 in.
Air and gas pressure.....	5.66 in.
Total pressure	12.00 in.

Such conditions are of course extreme, and would probably never be encountered; but in bad cases, it is possible for the net temperature drop to be cut in half. This would mean a reduction of 13° in the temperature of steam in the calandria (the net drop having been estimated at 26°): The pressure distribution in this case is as follows:

Vapor pressure corresponding to 156° F.....	8.76 in.
Air and gas pressure.....	3.24 in.
Total pressure, as previously.....	12.00 in.

The worst effect, and the one usually encountered, is the fact that this gas when present even in small quantities, forms an insulating film on the steam side of the surface which does not permit the free passage of heat, thereby greatly reducing the coefficient of transmission. It is quite evident that the trouble will be very much exaggerated if conditions of approximate quiescence obtain within the heating surface or any part thereof, and very much reduced if the reverse is true, for rapid vapor currents between the tubes tend to scour the insulating film, and bring about better heat transfer thereby.

This condition of air binding tends to be cumulative, for if there is a slowing up of the rate of evaporation, naturally, the film effect will grow worse, which will again react adversely, and so on.

The wise plan therefore is to make ample provision. In certain designs, all these gases are driven to one point by natural operating conditions, and their removal thus becomes very simple, necessitating only one connection, and permitting thorough venting with practically no loss of steam.

With other designs, it is necessary to install a number of pipes connected together with one control valve. Here it is advisable to make each individual connection relatively small, and to so arrange the piping that each pipe will draw its share. Lack of attention to this detail will result in one or two pipes venting excessively, while others will vent too little, or not at all. As to the location of the individual vent pipes in a given body, the best method is to spot them in zones of inaction, which can be located while the equipment is in operation, by lowering the levels, when the dead tubes can be identified by the fact that they do not spout as the others.

The main pipe from each body should be brought outside, where a

control valve must be provided. Between the control valve and the evaporator, a thermometer should be inserted, calibrated to indicate pressures and vacua as well as temperatures. The size of the pipes should be ample for facilitating starting up quickly, and in operation, should be throttled until the vacuum indicated on the thermometer in the air pipe is just a little higher than the vacuum in the steam belt being regulated. In this way, the losses of heat from this source will be held to a minimum. Such losses are never very considerable, and cannot be, unless the pipes are all out of proportion, for the specific volume of vapor is so large that it takes a great many cubic feet to make one pound.

As to the approximate setting of the valves in these vent pipes, according to practice, in the sextuple effect referred to previously, the air pipes out of each body would be 2 in., connected to a 4-in. header, leading to the vapor belt of the last effect, the vent on that particular body being independent. The valves should be globes, open for normal operation about as follows: First Effect, $\frac{1}{4}$ turn; Second Effect, $\frac{1}{4}$ turn; Third Effect, $\frac{1}{2}$ turn; Fourth Effect, $\frac{3}{4}$ turn; Fifth Effect, 1 turn; Sixth Effect, 2 turns. If the material discharged were all steam, it would only represent a fraction of one per cent of the amount actually used.

For vertical tube evaporators, it is wise to let the vent pipes attached to the top tube sheet extend downward to within not more than 4 in. or 5 in. of the lower end of the tubes, and to drill several holes near the top under the upper tube sheet for ammonia gas outlet. If the quantity of ammonia gas is considerable as in the beet sugar industry, it might be advisable to install two sets of vent pipes, one for light gases taking off in the top of the calandria, and the other extending to the bottom for air.

When starting all valves in the vent pipes should be wide open, and subsequently throttled, as suggested in the discussion on Starting Up. This will enable the operator to put the equipment in commission very much quicker. It is dangerous to open the vents from the pressure bodies too much, as their discharge will interfere with the venting of the other bodies, where the difference of pressure is not so great.

Chapter 8.

Feeding Systems.

There are many points worthy of careful consideration in the general methods employed to introduce the solution to be concentrated into the evaporator, and after its introduction, its proper removal and introduction into the next succeeding body, if the apparatus is a multiple effect. It is much more than a question of running liquor through a pipe from one container to another.

In the first place, the solution is to be concentrated while it is in the evaporator, and great care must therefore be taken not to dilute the outgoing liquor by admitting unconcentrated solution near the outlet. As more or less attention is given to the design, one of three conditions will exist in each body of the Multiple Effect:

(a) If the feed is poorly arranged, the outgoing liquor will be diluted by the incoming solution, and the average concentration in the evaporator will be greater than that of the outgoing liquor.

(b) In the average well designed equipment, the distribution of the incoming feed is such that there is a thorough mixture, and in this case, the concentration of the outgoing liquor is the same as the average concentration in the evaporator.

(c) By making special efforts, it is possible so to design the apparatus that the concentration is progressive in each body, entering at a certain density, and by following a predetermined path, gradually increasing to the final density at the outlet. Here, the outgoing liquor is greater in specific gravity than that of the average contents of the evaporator. This is called "Zone Concentration." It has been shown in other chapters that high concentration, and consequent increased viscosity have a very marked retarding effect on heat transmission. There can be no argument, therefore, as to the wisdom of adopting this method where the range of inlet and outlet densities permit a substantial gain. If, however, the change of density is very small in passing through the evaporator, the gain will be small in proportion. Usually, the great change in density occurs in the last body of a multiple effect, and at this point, it is possible to secure a very material improvement by making use of the zone concentration principle.

Another point to be carefully studied is the "flash" effect as the juice passes through the feed pipes of the evaporator from one body to the next. Here, inasmuch as the liquid goes from higher to a lower zone of pressure, a flash will occur, corresponding to the drop in temperature.

Taking for an example the operating conditions given for the sextuple effect whose heat balance was elaborated in another chapter, the following table gives for each body the liquor passing through the feed pipes, the drop in temperature of this liquor, the B.t.u.'s released by the flash, the value of latent heat L for each set of conditions, the vacuum or pressure, the specific volume of steam corresponding, the actual flash evaporation, in weight and volume, the volume of liquor passing through in cubic feet per hour, the total volume of vapor and liquor in cubic feet per hour, the same in cubic feet per second.

DATA FOR DESIGN OF FEED PIPES OF SEXTUPLE EFFECT.

(See Chapter Practical Use Heat Balance.)

Bodies	Liquor per Hr.	Temperature Drop	Released B.t.u.	Latent Heat	Evaporation Corresp.	Vacuum or Pres.
1.....	100,000	..				
2.....	90,100	9	810,900	967.2	838	1.5 lbs. V.
3.....	79,400	11	873,400	974.4	895	4-in. V.
4.....	67,800	14	950,000	983.6	965	11-in. V.
5.....	55,100	20	1,102,000	996.4	1,105	18-in. V.
6.....	41,000	40	1,640,000	1,021.6	1,605	26-in. V.

Bodies	Spec. Vol. Vapor	Flash Cu. Ft./Hr.	Liquor Cu. Ft./Hr.	Total Cu. Ft./Hr.	Total Cu. Ft./Sec.	Liquor Cu. Ft./Sec.
1.....						
2.....	24.4	19,600	1,360	20,960	5.81	0.378
3.....	30.3	27,100	1,180	28,280	7.83	0.328
4.....	40.5	39,100	995	40,095	11.13	0.277
5.....	78.9	87,200	790	87,990	24.50	0.220
6.....	178.4	286,500	564	287,064	79.70	0.157

Doubtless, the data contained in this table will prove very surprising to those who have never made a set of such calculations, and show definitely why it is necessary to use such large pipes to feed Multiple Effects. As can readily be seen, the determining factor is the volume of the flash, and not so much the volume of the liquor. The flash is always many times the volume of the liquor producing it. In the case of the last effect in the selected example, this flash is about five hundred times the volume of the liquor.

Conditions as described above will obtain particularly when feeding from the bottom of one evaporator to the top of the succeeding one. In reality, the effect is not instantaneous, and, of course, the extent of the flash will depend upon the pressure existing at the particular point at which it takes place. When feeding into the bottom of the body instead of the top, the flash will be reduced very materially. Thus for the above example, if the juice is released at a point in the last effect 3 feet below the static level of the liquid, the following conditions will exist:

The specific gravity of the liquor is 1.29, therefore the vacuum at this point will be reduced by

$$3 \times 1.29 \times \frac{30}{34.5} = 3.36 \text{ in. of mercury,}$$

3 represents the static head,

1.29 represents the specific gravity of the liquor,

30 represents the barometric column in inches of mercury,

34.5 represents the barometric column in feet of water.

In other words, instead of flashing at 26 in. vacuum, the reaction will take place at $26.00 - 3.36 = 22.64$ in., or at a temperature of 24° below, so that instead of having a 40° flash as shown in the table, there will be only 16° , making the liberated heat 655,000 B.t.u.'s per hour. L at 22.64 in. vac. = 1008, and the corresponding evaporation from this flash will therefore be $\frac{655,000}{1008} = 650$ lbs. per hour. The specific volume

of steam at 22.64 in. vac. is 99.2 cu. ft. per pound. The volume of the flash at this point will therefore be $650 \times 99.2 = 64,500$ cu. ft. per hour, or 17.9 cu. ft. per second. The volume of the liquor is the same, 0.157 cu. ft. per second, so that the total discharge through the feed pipe will be $17.9 \times 0.157 = 18.057$ cu. ft. per second, instead of 79.7 as before.

It seems that the logical way to overcome this problem is to enlarge the cross section of the feed pipe on its discharge end, which is a method used successfully for many years. In fact, it has been good practice to utilize the heat liberated by the feed flash by distributing it very thoroughly in the bottom of the evaporator, thereby preheating the liquor that has come down through the downtake before it reenters the tubes. In view of the hydrostatic pressure above the point of feed entry, by the use of a proper "silent heater" with confined circulation, this can be done very nicely, and it results in a material benefit to the heat transmission besides practically doing away with the vibration which would otherwise occur when the flash is condensed. Since the boiling temperature at the point of entry is very much higher than the corresponding liquor temperature above the top tube sheet, there will be no difficulty from this source, unless the volume of liquor in actual circulation is very small, conditions very unlikely except in long tube evaporators when the boiling levels are too low.

If the liquor is admitted into the bottom without proper distribution, the enormous volume of flash will be local, and will cause the tubes immediately above to spout violently, with consequent risk of entrainment losses. Sometimes, there is also serious pounding which sets up dangerous vibrations in the apparatus and its supporting platform.

In arranging the distribution of feed in the bottoms of evaporators, pipes with small perforations must be avoided, as these small holes soon plug up with scale which is loosened during the period of cleaning. No hole should be less than 1 in. diameter, and it would be better to have them $1\frac{1}{2}$ in. diameter.

It is seen, therefore, that these matters should receive the careful attention they deserve in order to keep out of difficulties, and take advantage of conditions favoring successful operation of the apparatus.

In another chapter (Circulation) were mentioned the problems en-

countered when the liquid entering the evaporator is substantially cooler than the temperature of ebullition. In this case, as has been carefully brought out, it is advisable to spray the cold liquid into the vapor space, thereby heating it by direct contact with vapor. This will give very much better results than feeding cold solution below the bottom tube sheet, as such an operation involves a very inefficient heat transmission at the bottom of the tubes on account of the sluggish circulation existing at this point. It is not inappropriate to reiterate the fact that evaporator surface is designed for evaporation only and not for heating, and that it will yield very poor results when such a reaction is attempted.

While on this subject it is also well to call attention to the axiom that best results are obtained by continuous operation and uniformity of work. It is bad practice to feed intermittently, as it upsets the working balance of the equipment. The feed valves must be throttled to permit a continuous and uninterrupted flow at normal working rates, maintaining uniform levels.

This factor is so important that it has led to the development and practical use of automatic feed systems which control these variables independently of the operator. Some of these will be described in the following pages.

Hand Control—Bottom to Top Feed. Fig. 1 shows the simplest and most common arrangement of feed piping for multiple effect operation. This is probably the oldest system in use.

The liquor being concentrated is admitted above the top tube sheet of the first effect, and removed from the bottom at the centre, whence it

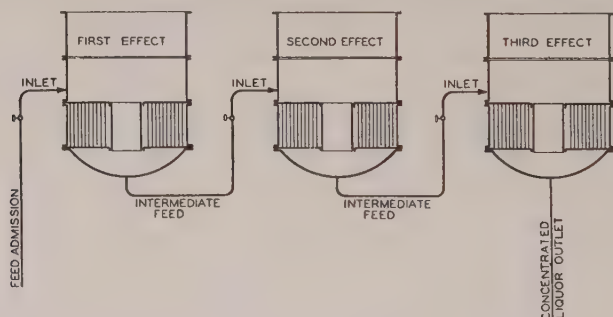


FIG. 1.—Hand Control System. Bottom to Top Feed.

is transferred to the next succeeding body, the admission being again at a point above the top tube sheet, and so on through the succeeding bodies, the final removal being from the bottom of the last body.

The outstanding advantage of this arrangement is its simplicity and low cost, and there are a great many evaporators to-day so arranged, particularly those of European manufacture. The flash from the feed pipe, taking place in a horizontal direction, does not impose any serious danger of entrainment. Also, inasmuch as the incoming hot liquor enters above the

liquid level in the evaporator, there will be neither pounding nor vibrations from this source, and the operation proceeds quietly and without disturbance. From the operator's standpoint, it is quite satisfactory.

There are a number of weak points which it is well to bring out. First, after washouts, if the solution being concentrated is of a scale-forming nature, trouble may be experienced from clogging of the intermediate feed pipes with scale which always drops from the tubes when the apparatus is put into commission immediately after a cleaning operation. This tendency can be overcome to a great extent by placing a well, or scale trap at the bottom of each body, which acts as a depositing chamber, storing this scale until the next washout period.

Another point is that the fresh low density feed is admitted into the liquor after it has passed through the tubes, where concentration has taken place, and thus a partially diluted liquor goes down the downtake, and it is this mixture which is withdrawn from the evaporator. Unless the ratio of circulation is fairly great, the density of the outgoing liquor will therefore be less than that of the liquor as it leaves the tops of the tubes, a defect mentioned previously.

Particularly if the tubes are long, the feed pipes will have to be very large in proportion to the quantity of liquor passing through, for the flash will have the full volume, as indicated in the table, and the discussion at the beginning of this chapter.

It must be remembered that the cause of flow of liquor through the feed pipes of evaporators is the difference of pressure or vacuum between adjacent bodies of multiple effects. If the difference of pressure is small, such as is necessarily the case as the number of bodies in the set is increased, then it is necessary to use lower juice velocities, and to arrange the piping so as not to rob any of this difference of pressure. In the feeding system indicated in Fig. 1, this factor has been overlooked.

To illustrate, presuming that the tubes are 60 in. long, that the operating level is 20 in. above the bottom tube sheet, and that the feed inlet point is 8 in. above the top tube sheet, it will be seen at once that in addition to passing the liquor from one body to the next, it must also be lifted from the level existing in one body, 20 in. above the bottom tube sheet, to the level of the liquor inlet in the succeeding body, a vertical distance of 48 in., which of necessity must encroach upon the working pressure difference between the two bodies under consideration.

Again, referring to the specific example at the beginning of this chapter, below is given a table in which will be found the information bearing on the subject:

Bodies	P. or V.	Pres. Dif. In. Hg.	Height, Lev. to Inlet	Sp. Gr. Liquor	Cor. Head In. Hg.	Net Dif. In. Hg.
1.....	5.0 lbs.	7.0 in.	4.0 ft.	1.065	3.70 in.	3.30 in.
2.....	1.5 lbs.	7.0 in.	4.0 ft.	1.079	3.75 in.	3.25 in.
3.....	4.0 in.	7.0 in.	4.0 ft.	1.092	3.80 in.	3.20 in.
4.....	11.0 in.	7.0 in.	4.0 ft.	1.120	3.90 in.	3.10 in.
5.....	18.0 in.	8.0 in.	4.0 ft.	1.167	4.05 in.	3.95 in.
6.....	26.0 in.					

Thus it will be seen that the actual pressure difference between adjacent bodies is cut in half, which means that slower velocities must be used, or it will become impossible to transfer the feed.

As to the actual liquor speeds permissible under these conditions, by reference to the chapter on Circulation, a clear idea will be obtained of the problem involved. Roughly speaking, there is a net pressure difference for feeding of about 3.4 in. of mercury on the average, which corresponds to about 3.91 ft., which in turn is equivalent to 47 in. head. Under these conditions, such a head would permit the following combinations:

Inlet Liquor Velocity	Velocity Mixed Vapor and Liquor
5.0 ft. per sec.	32.0 ft. per sec.
4.0 " " "	40.0 " " "
3.0 " " "	52.0 " " "
2.0 " " "	73.0 " " "
1.0 " " "	115.0 " " "

The above figures do not make any allowance for friction, which must be taken into account, and, therefore, these speeds must be discounted somewhat. Thus it can readily be understood that trouble may be expected in feeding the last effect with this system, unless ample allowance has been made, and as suggested before, much is to be gained by enlarging the feed pipe on the discharge end, particularly within the flash zone, which is easily calculated when the existing vacua and pressures are known.

Hand Control—Bottom to Bottom Feed. The feeding system shown in Fig. 2 is a hand control arrangement in extensive use in the cane sugar

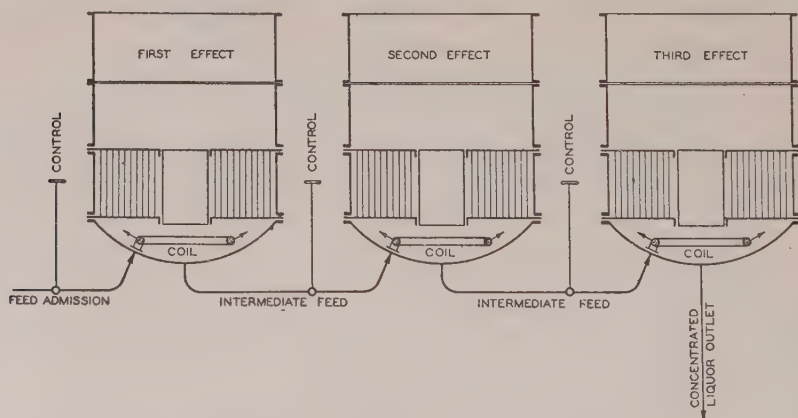


FIG. 2.—Hand Control System. Bottom to Bottom Feed.

industry. Here, the feed is admitted into the bottom of each effect by means of a closed coil having large perforations on its outer periphery only. The area of these holes should be three or four times the net area of the feed pipe.

By this arrangement, the weak liquor is thrown towards the outer circumference of the evaporator, the flash absorbed by the cooler circulating liquor, this mixture resulting in a dilution for the outer tubes of the evaporator. The concentrated liquor returning through the downtake is undiluted until in its eccentric motion, it passes the feed coil. In this way, the concentrated liquor withdrawn from the bottom at the centre, is at the maximum concentration in the evaporator.

It is wise, as suggested in the previous system, to put in a scale trap in order to keep the feed pipes clear, and this precaution is really doubly important in view of the fact that the discharge to the next body is through a perforated coil, which would undoubtedly clog much more easily than the open pipe in the other case.

As to the size of the feed pipe, in view of the fact that the liquor is transferred from bottom to bottom, the net difference of pressure or vacuum between adjacent bodies is not encroached upon as in the case of the bottom to top feed discussed previously, and normal liquor velocities are permissible. It has been found in practice that the upper limit of such liquor velocities is about 5 feet per second, and it is better to use $3\frac{1}{2}$ to 4 feet as a working speed, which provides leeway for throttling the feed valves. A properly calculated heat balance will show the quantities passing through each pipe, and no difficulty will be encountered if the piping is designed within the limits specified above.

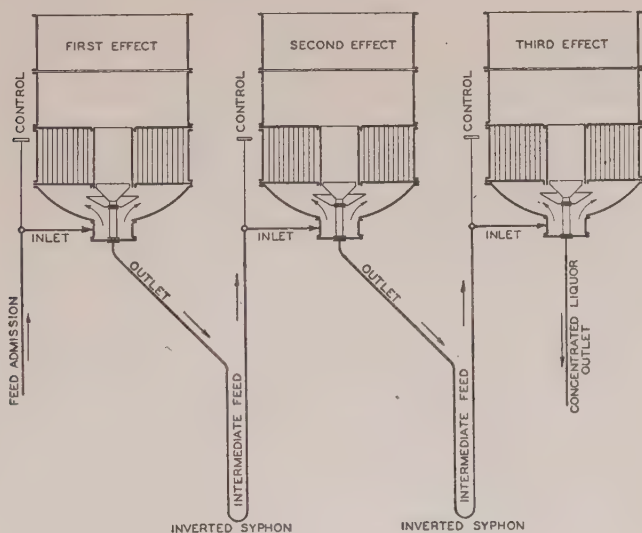


FIG. 3.—Chapman Feed Control System.

The Chapman Circulator. The Chapman Circulator, shown in Fig. 3, is probably the pioneer in the field of automatic feed controls. It has been successfully installed in many plants.

As indicated in the sketch, the liquor is fed into the bottom well, above which is placed an inverted cone, the object being to deflect the juice which makes only one pass through the tubes, and after reaching the top tube sheet, flows to the downtake, which, in this case, is closed at the bottom with a cone from the lower part of which the liquor is withdrawn.

By referring to the chapter on Natural Circulation, it will be understood that the velocity of the liquid as it enters the tubes from the bottom, is very sluggish except where locally affected by the improperly distributed flash. The volume of liquor coming from the feed pipe must be distributed throughout all of the tubes, and therefore, if the speed here is only 4 or 5 feet per second, when this is spread out to supply the entire cross section of the tubes, the entrance velocity must of necessity dwindle down to practically nothing. It is not possible to avoid local concentration in some of the tubes according to location, to a point far in excess of the outlet density.

When in operation, the appearance is that of having levels entirely too low, and some of the tubes appear to be dry. Some manufacturers have modified this design by installing additional downtakes, but it is evident that the inherent defect of too low levels cannot be overcome. Be that as it may, there are many Chapman controls in successful operation, and as such, they must be recognized.

This arrangement brings us to the consideration and study of the so-called inverted syphon, or U-pipe. The feed, as it leaves the bottom of each evaporator, goes down a pipe, as indicated, to a considerable depth, and then comes back, discharging into the next body. The object of this loop, or U-pipe, is to form a seal against the difference of pressure between the two adjacent bodies. The presumption is that the level of the liquor in the outgoing leg will be such that the hydrostatic head between this level and the level of the liquor in the body into which liquor is being discharged will balance the difference of pressure or vacuum between the two effects under consideration. This, of course, is on the assumption of static conditions. When the liquor is flowing through, the level on the inlet end rises enough to provide sufficient extra head to cause a flow to take care of the quantity of liquid entering the syphon, thus overcoming friction, acceleration, flash head, etc.

Inverted syphons, particularly if the pipe is large in proportion to the quantity going through, will oscillate, sometimes violently, and it is necessary to make the depth such that these oscillations will not pass the bottom of the pipe, thus breaking the seal, which would probably cause a water hammer, with danger of breakage.

It is customary to put a valve on the upper end of the upflow leg, and throttling this valve produces the same effect as though the U-pipe has been lengthened within certain limits. It is good practice to use a regulating check valve at this point, which eliminates the oscillations, and thus avoids all difficulties from this source. Under such conditions, if the length of the syphon is sufficient to provide a maximum difference of head 50 per cent. in excess of the expected pressure difference between the two bodies, no difficulty whatever will be encountered.

It must be noted that as the liquor approaches the discharge end of the syphon, a flash will begin, when the hydrostatic pressure superimposed upon the pressure on the discharge side is less than that at the inlet side, a condition which will generally be encountered below the control valve when located as shown on the sketch. When this flash is serious, it causes certain vibrations, and the syphons should therefore be steadied or supported at their lower ends.

Webre Automatic Level Controls. Fig. 4 shows a system used for controlling levels in evaporators. The scheme is based on bottom to bottom feed as in Fig. 2, with a scale trap to keep the piping clear.

There is provided on the outside of the evaporator a fair sized chamber

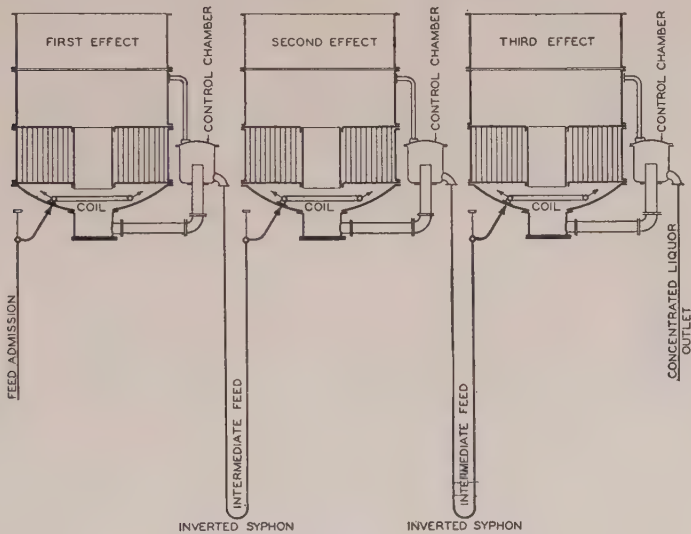


FIG. 4.—Webre Automatic Level Control Outside Overflow.

connected to the bottom by means of a large pipe. This chamber has an equalizing pipe to the vapor space above. The liquor, coming from the bottom, overflows the large pipe in the centre of the chamber, thereby controlling the level in the evaporator at that point. This liquor, after dropping into the bottom of the control, is removed and transferred to the next body by means of an inverted syphon similar in all respects to that described in Fig. 3. The overflow pipe is made adjustable either from the inside or from the outside, and the chamber is provided with a sight glass which indicates the operation.

This equipment is not subject to the objections of that shown in Fig. 3, and has given very good results in practice. The liquor inlet to the chamber should be roughly twice the diameter of the feed pipe, and the equalizing pipe should be ample, about one-half the feed pipe diameter.

Inside Overflow Control. Fig. 5 shows the inside overflow method of level control. There are a number of such arrangements in use. In their operation, they are very similar to the Chapman circulator.

On account of the position of the funnel, which catches a large part of the liquid as it drops in the downtake, the level may be considerably lower than that of the funnel top. The position of this overflow is usually set in the field permanently to accommodate operating conditions.

Generally, the feed is admitted into an external coil as shown, having four branches connected to the bottom. The distribution of the feed and flash absorption are not nearly so good as with the internal coil. There is a practical advantage, however, in keeping the bottom clear of obstructions,

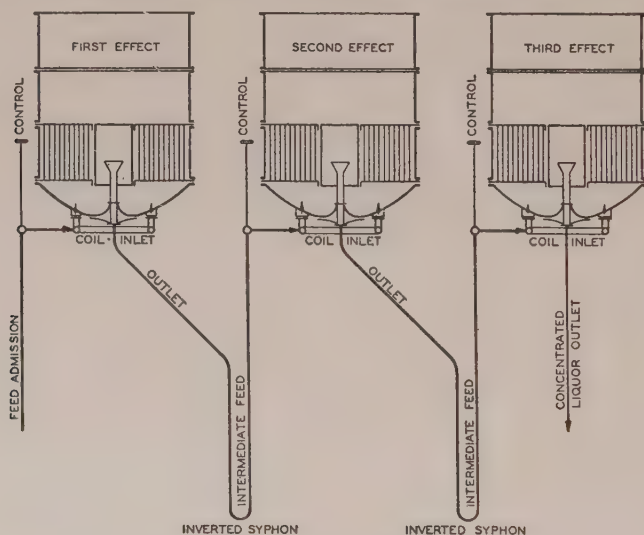


FIG. 5.—Inside Overflow Level Control.

thus facilitating repairs. Usually, there is a screened pocket above each coil branch to distribute the feed. The piping on account of its arrangement, suffers very little from scale choking. The U-pipes are similar to those in Fig. 3.

Float Level Control. In Fig. 6 is shown an arrangement for automatically controlling the levels by means of float valves. In operation, the results are similar to those obtained by Fig. 4.

The float chambers are closed, and equalized with the bottoms and tops of the evaporator. The vertical motion of the float is transferred by means of a lever to a crank passing through a stuffing box on the outside, which in turn is transferred to a butterfly valve or some such device in the main feed line, thereby controlling the flow of liquor so as to maintain the level as determined by the float.

In this scheme, the liquor is pumped out of the last body at such a rate

that the ultimate density remains at the required point, and when it works, its operation is simple and satisfactory. It will be noticed that this is the only scheme controlling by the inlet. All others control by the outlet. The main troubles are in the floats, which tend to become water logged; and the stuffing boxes, which bind due to incrustations and liquor leaks.

This arrangement or modifications of it has been used to a certain

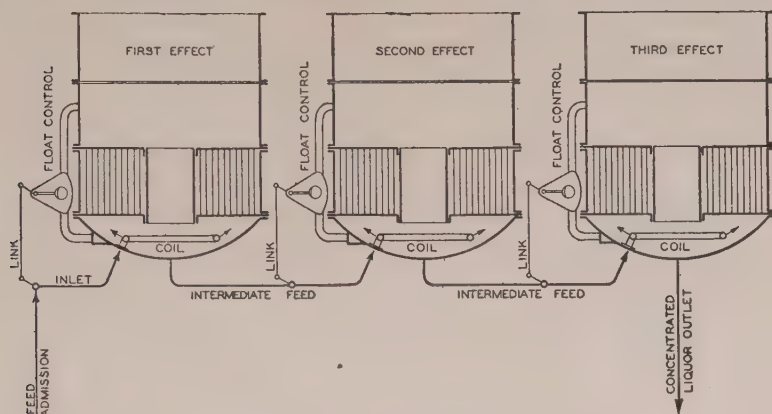


FIG. 6.—Automatic Level Control with Float Valve.

extent, but the prejudice against floats and movable parts has prevented its extensive adoption. Such objections are well taken, and it is doubtful if they can be satisfactorily overcome, when it is possible to secure as good if not better results by other and simpler means.

Circulation Controls. In Fig. 7 is shown the system designed by the author for feeding evaporators to conform with all the points brought out in theory and practice of multiple effect operation. This scheme uses the top to top feed with froth overflow.

Here, no attempt is made to regulate the levels of the liquid as represented by the liquor level gauge. What is controlled is the level of the froth or liquor in ebullition above the top tube sheet. This in general determines the actual amount of liquor in circulation up the tubes and down the downtakes.

Each body is divided up by means of a copper diaphragm placed upon the upper tube sheet, thus making up to zones of concentration, the inlet zone, enclosing about two-thirds of the heating surface, and the outlet zone, enclosing about one-third. The tube sheets are equipped with distributed downtakes, so that there is circulation in both zones. In order for the liquor to pass from the inlet to the outlet zones, it must first circulate through the tubes and downtakes of the former, pass under the bottom tube sheet, proceed upwards through the tubes, and leave above the upper tube sheet.

The feed pipes entering the evaporator are enlarged greatly to take

care of the flash which occurs in a generally horizontal direction, and at a relatively slow velocity in view of the large inlet. There is no difficulty from scaling of feed pipes, as the concentrated liquor is removed from the top after the scale has dropped out in the bottom.

No adjustments are necessary, and the levels will vary according to working conditions always settling to the proper point for the predetermined circulation, irrespective of the imposed conditions of operation, condition of the heating surface, or rate of evaporation. There are no internal connections for either feed inlet or outlet, and the scheme as a whole is very simple.

This so-called Circulation Control has been used under most trying

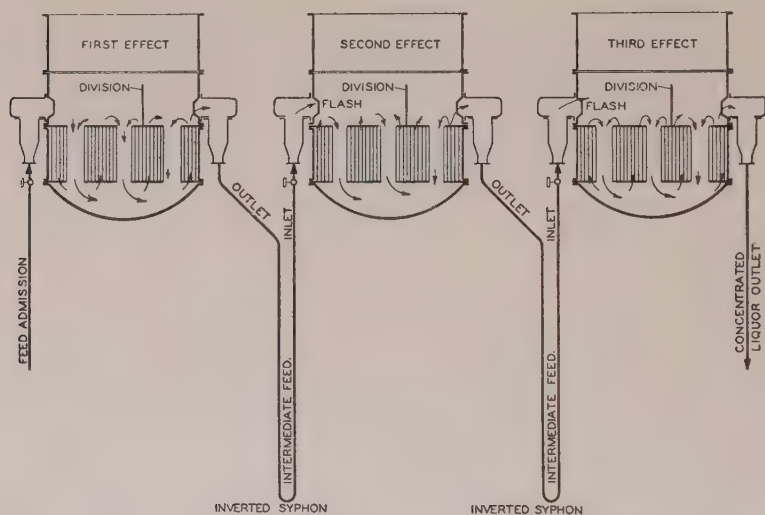


FIG. 7.—Webre Circulation Control with Distributed Downtakes and Zone Concentration.

conditions with good results. The U-pipes are proportioned along the same lines as in the other systems discussed previously.

Special Cases. The above discussion of feeding systems applies to parallel current operation of multiple effects. For counter current operation, which finds only few applications, it is necessary to pump the feed back from each body to the preceding one, which is in reality a very simple operation hardly worthy of elaboration here. It is necessary to know the quantity of liquor to be transferred, its specific gravity, and the head against which it is to be pumped, represented by the difference of pressure between adjacent bodies, to which must be added friction.

In view of the fact that the feed temperature will generally be lower than the temperature of ebullition, it is advisable to feed above the top tube sheet in all cases, spraying into the vapor space, these conditions being similar to first body operation with cold feed, which has already been dis-

cussed. In the case of parallel feed operation, this is very simple, and is so seldom used that it needs no further mention here.

Batch Concentration. In all of the above cases, it has been assumed that the solution to be concentrated was being fed continuously into the first body of the apparatus, and removed continuously from the last body.

In some instances, when the ratio of the evaporation is very great, as in the case of the concentration of wood extracts, where it sometimes takes 40 gallons of solution to make 1 gallon of final product, it pays to operate on a different scheme. Here it is customary to run the evaporator "dead end," or without withdrawing anything out of the last effect, allowing the concentration to go on until the required density is attained, when the apparatus is stopped, and the concentrated liquor in the last body is dumped. When such a procedure is employed, it is better to have a large liquor capacity to the last body by providing a deep bottom, and in this way, the dumping periods are put further apart, with less loss of time. It is not necessary to break vacuum to make the "strike." Simply shut off steam, and remove the contents of the last body by using a concentrated liquor pump properly piped to operate under vacuum. (See chapter on Condensate Removal.)

After this operation, the last effect is refilled with raw liquor, and the operation resumed as before.

The advantages of this method of operation under these special conditions is that in the first place, the equipment does not foul nearly so rapidly. Also a more accurate control of the density can be obtained than by the other method of continuous concentration, as this would involve an almost microscopic adjustment of the valves, inasmuch as such a small quantity of liquor must be discharged. There is another distinct advantage, and it is that by this method, the heating surface of the evaporator in the last effect does not have to overcome the constant high density which would otherwise exist, except for a short period at the end of the "strike," and therefore can transmit heat very much faster and better at all times except this small period.

Continuous Batch Concentration. There is another method of operating the batch system without interrupting the work of the evaporator, as shown in Fig. 8. Here, a batch tank is used into which raw liquor is fed by a float controlled valve to keep the level in this tank at a fixed point. This liquor is pumped from the batch tank through the evaporator, and back into the tank again, the quantity thus circulated being about 50 per cent. more than it would be for continuous concentration. Thus the density in the tank slowly builds up, its level remaining the same. When the density coming out of the last body is at the point of final concentration, valve "A" is closed, valve "C" is opened, and valve "B" is closed. Then, the entire contents of the batch tank are pumped out through the evaporator, leaving it at the proper density, and going to the finished storage.

When the tank is empty, valve "A" is opened, and when the liquor density leaving the last body begins to fall, valve "B" is opened, and valve "C" closed, and thus a new batch is started. In the meantime, the evap-

erator has not been stopped, but the density in transit has been changed. This method is practically impossible with level controls on account of the wide variation of density and rates of evaporation during the cycle. It does work beautifully, however, with the circulation controls as shown in Fig. 7, for which they were designed.

Due to the wide range of variations to the density, the heating surfaces do not foul as rapidly as otherwise. The operation is very simple, and

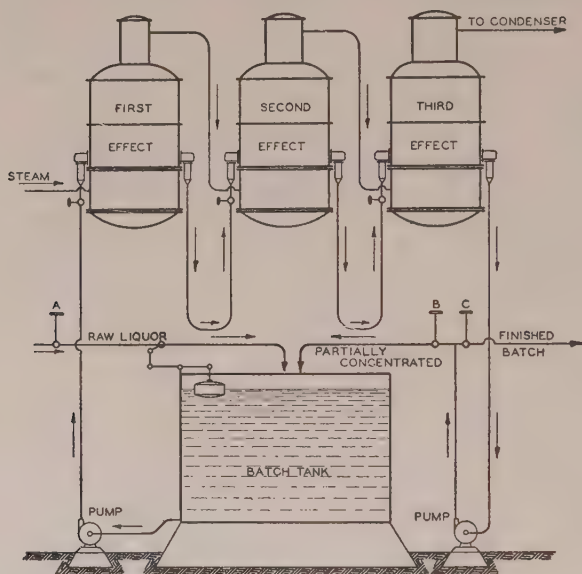


FIG. 8.—Continuous Batch System.

requires practically no attention except when pumping out a finished batch. One large plant running on this scheme, pumps out for about 30 minutes every 6 hours. This period depends entirely upon the size of the batch tank, and can therefore be arranged at will.

One caution is necessary. The admission of raw and partially concentrated liquor into the batch tank must be so arranged that the two will mix thoroughly, and for this reason it is better to have a cylindrical tank, and to admit the liquor on a tangent, thereby giving the mass a swirling motion to prevent stratification of the heavier liquor, and upsetting of the process.

Chapter 9.

Practical Operation of Jet Condensers.

In the theoretical part of this book, the subject of jet condensers has been discussed somewhat at length. The practical operating side of these equipments will now be considered.

The first essential is the water supply. The quantity of injection necessary will depend first upon the amount of steam to be condensed, and the vacuum to be maintained. It will also depend upon the initial temperature of the water.

It has been mentioned in the chapter on condensers that the most important function is the heating of the injection water to as near the temperature of the incoming steam as possible. It is found in actual practice that the outlet temperature of this water will be from 5° to 25° cooler than the steam temperature. Perhaps, a fair average to be expected is 15°, with good equipment. The weight of water per pound of steam condensed, or what is called the ratio of injection, will vary according to the following formula:

$$W = \frac{L + (T_3 - T_2)}{T_2 - T_1}$$

Where W = Pounds of water per pound of steam condensed,
 L = Latent heat of steam at its temperature and vacuum,
 T_1 = Temperature of water entering condenser, ° F.,
 T_2 = Temperature of water leaving condenser,
 T_3 = Temperature of steam entering condenser.

Thus with 26 in. vacuum, the temperature of steam is 125, T_3 ; L is 1021.6; assume the injection water at 70 for T_1 , and the water out of the condenser 15° cooler, or 125 — 15 = 110, T_2 ; by substitution in the above formula, one would have:

$$W = \frac{1021.6 + (125 - 110 = 15)}{110 - 70 = 40} = 25.91.$$

In other words, it would take practically 26 lbs. of injection water to condense 1 lb. of steam. For condensing 20,000 lbs. of steam per hour, the injection required in gallons per minute would be:

$$\text{G.P.M.} = \frac{20,000 \times 26}{8.33 \times 60} = 1040.$$

In estimating the water requirements, it is necessary to assume worst seasonal conditions as to temperatures, and the minimum vacuum permissible under such conditions. One must not be scant with the water supply, and a liberal margin must be provided over and above the actual maximum theoretical requirements.

In selecting the pumps for this service, sufficient head must be provided to deliver water to the condensers at atmospheric pressure. This is particularly true if there are several condensers, and indispensable if the condensers are of the ejector type, for these will not start without a fair water supply. Very often, this precaution is overlooked, and for the purpose of saving a little power, the vacuum in the condenser is relied upon to decrease the head on the pumps to within their range. This is very poor economy, involving the risk of unsatisfactory and unreliable operation.

There should be at the injection valve a sufficient difference of pressure to permit throttling, thereby making it possible properly to distribute the water supply to the various condensers. If this difference does not exist, it generally happens that one condenser robs the others, thus upsetting the vacua in the various units, and this defect will be particularly aggravated when starting up a unit that has been shut down.

A very wise precaution, when there are a number of condensers, is to put in two valves in each injection pipe, one of which is set at the right opening, determined by trial, and locked in position. The operator is then permitted to use the other valve for adjustment or cut off. In this way, there will be no upsetting.

In selecting the water pumps for this service, the impellers should be such that if the head is reduced below the point for which they were designed, the motors will not be overloaded. This, of course, applies to motor-driven centrifugal pumps, which are in common use for this purpose. Many motors have been burned out due to lack of attention to this feature.

Sometimes, the problem of water distribution is attacked from another direction by providing an elevated overhead open top tank, which is kept full by the pumps, and from this tank each condenser draws its water by means of its own individual pipe, thus avoiding interference. If the tank is made in the form of a stand-pipe, extending below, with the connections for condensers at the bottom of the tank, should the pumps prove to be insufficient to maintain the proper levels through breakdowns or interruptions, the level in this tank drops, decreasing the difference of head between the condensers and the tank, until the flow of water through the pipes, determined by this head, slows down to correspond to the altered capacity of the pumps. In the event of a complete pump shutdown, the level in the tank will drop further, until a barometric seal is established, corresponding to the then existing maximum vacuum in the condensers, and thus a sudden break of vacuum through the injection pipes drawing air is avoided. In some industries, such as the manufacture of sugar, this feature is very important, particularly with the vacuum pans.

In arranging the pump installations, the suctions should be protected

with screens to avoid clogging the runners of the pumps with extraneous material, and this is especially important if the condensers have injection nozzles, or perforated spray pipes. It is needless to state that the piping should be designed of proper size for the quantity of water, and that ample allowance must be made for friction, velocity head and entrance head, information for which can be found in handbooks. It is also advisable in arranging the piping to avoid high points which invariably cause air pockets, thus materially interfering with the unrestricted flow of water.

As to the matter of the size of the pumping units, the preference is for three pumps, two of which are capable of carrying the load and the other a spare. Reliability in water supply to the condensers is of prime importance, and this subject is therefore worthy of careful consideration, for failure at this station means shutting down the plant, with all that this entails.

There has also been discussed in the chapter referred to previously the matter of air removal. The amount of displacement thus required is usually reckoned upon the amount of air in solution in the injection water, which is a function of its temperature. (See tables for air in solution at various temperatures.)

Taking for elaboration the condenser example used previously, the injection was 1040 gallons per minute, or 133.6 cu. ft. At 70° F., the air in solution is 3 per cent. by volume, thus giving 4 cu. ft. of free air per minute to be removed. In the condenser, this is very much expanded. The assumed vacuum, being 26 in., and the barometer 30 in., the absolute pressure is therefore 4 in. Assuming that the condenser is of the counter current type, in which air may be expected to leave at a temperature roughly 20° above the water temperature, or in this case $70 + 20 = 90$. The vapor pressure corresponding to 90° is 1.417, and the air being vapor saturated, one will have the following conditions:

Total absolute pressure in condenser.....	4.000 in.
Partial vapor pressure corresponding to 90°.....	1.417 in.
Residual air pressure.....	2.583 in.

The volume of the air must therefore be expanded from atmospheric pressure, 30 in., down to 2.583 in., and the volume per minute in cubic feet will be:

$$\text{Volume} = \frac{4.00 \times 30}{2.58} = 46.6 \text{ cu. ft.}$$

This amount is doubled to take care of air leakage in the condenser and apparatus preceding it, thus making the required displacement 93.2 cu. ft. per minute. The volumetric efficiency of the ordinary vacuum pump is only about 60 per cent. at that vacuum, so that the piston displacement required will be $\frac{93.2}{.60} = 153.5$.

While on the subject, it is well to call attention to a very interesting

fact about vacuum pumps. The maximum power will be required when the vacuum is in the neighborhood of 18 in. This power is less above and below this point, and when raising vacuum on an installation, if the pump is insufficiently powered, it will stop before the vacuum reaches this point. When this difficulty is encountered, it is possible to get over the hump by closing off the suction valve quickly and completely, subsequently opening it very slowly, thus keeping within the available power. Thus when the full vacuum is reached, the valve can be opened wide, and normal operation resumed.

This feature is very puzzling to the layman, but is nevertheless true, as an examination of the air cards of vacuum pumps will show the maxi-

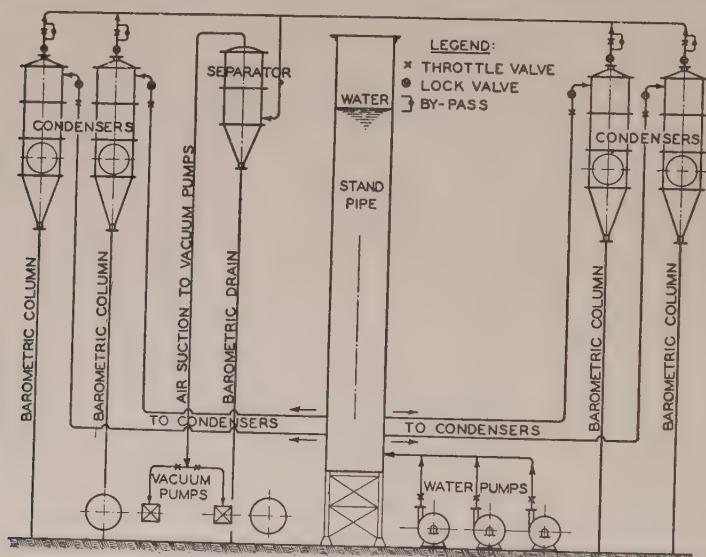


FIG. 1.—Condensing Plant with Water and Vacuum Systems.

imum mean effective pressure at about 18 in. vacuum, with reductions in either direction. The mean effective pressure with a blanked suction is practically zero. This explains why it is possible to resort to the procedure mentioned above. It must be stated that this is only possible with a fly wheel pump.

In piping up vacuum from the condensers to the pump, a trap or recipient should be installed above, with an ample barometric drain, for the thorough separation and removal of water which is entrained from the condenser, otherwise there is danger of wrecking the air end of the pump with a gulp of water if it is of the rotative dry type, and of high piston speed.

Fig. 1 shows an installation of four condensers arranged according to the discussion in this chapter, and needs no further explanation.

The reader has already been cautioned, in the other discussion, against installing long and tortuous tail pipes or barometric columns in condensing plants, and to make the descent as direct as possible.

Usually, these tail pipes are entirely too large. There is no necessity of using water velocities smaller than 7 to 10 ft. per sec. In fact, a large tail pipe is a decided disadvantage, for by maintaining a relatively high velocity, a considerable proportion of air and non-condensable gas is carried down and discharged into the hot well, thus relieving the load on the vacuum pump, a feature well worthy of elaboration. If the velocity in these pipes is too sluggish, no air whatever will be entrained, thus placing the entire load on the vacuum pump, and besides, permitting the bottom of the condenser to become partially air bound, thus decreasing its water heating efficiency.

In practice, it must be remembered that increasing the vacuum pump displacement to abnormal proportions will not necessarily increase the vacuum, which is limited by the temperature and quantity of water admitted to the condenser. It is also true that if the vacuum pump capacity is insufficient, increasing the quantity of water beyond a certain point will not only not raise the vacuum, but may actually reduce it. The two factors must be considered jointly, and not separately. Each has its well defined functions, and when these are accomplished, it is impossible to go further, and it is useless to build up hopes of improved operation on faulty premises which violate elementary principles.

If the air coming out of the condenser and going to the vacuum pump is relatively cold, it is possible to improve the vacuum by additional pump displacement, providing there is also a large difference between the temperature of the tail pipe water, and that of the vapor entering the condenser—otherwise, not.

If the air coming out of the condenser is hot, and the temperature of the water leaving the condenser is near that of the entering steam, it is possible to increase the vacuum by putting on more water, and conversely. As in all these things, practice and experience will be a good guide.

When several condensers are evacuated by one vacuum pump system, it is necessary to have throttle and lock-set valves in the vacuum pipe of each, and these valves must be adjusted so that the temperature of the air leaving each condenser is substantially the same. Once the set valves are properly adjusted, they should be locked in place, allowing the operator to adjust the throttle valve to suit his conditions.

When cutting in a new condenser, great care must be exercised in order to do it gradually, otherwise, the vacuum at all other points of the system will drop. This is best accomplished by the use of a small by-pass across the throttle valve.

A better scheme, where conditions permit, is to have a separate vacuum pump to evacuate the air only, after which the unit is cut into the vacuum system, thus avoiding these disturbances. This is particularly advisable in the sugar industry, where the various vacuum pans strike at regular intervals, and have to be evacuated for a new batch three or four times a day. Here, the vacuum in the operating pans must not drop

below $23\frac{1}{2}$ in., otherwise, it will interfere with the formation of grain.

As suggested in the other discussion of condensers, these troubles are greatly minimized by using ejector condensers which do not require vacuum pumps, and whose vacua are therefore independent of one another.

It is very annoying with central vacuum systems to locate leaky units, and when such an emergency arises, all of the condensers suffer from the fault of the leaky one. It is possible, by carefully observing the air outlet temperature of each condenser to single out the defective one, for in view of the larger amount of air going through the vacuum pipe, its temperature should be sensibly lower than that of the others, if all the valves were properly set in the beginning. It is nevertheless quite an annoyance.

Chapter 10.

Vacuum Pumps and Their Characteristics.

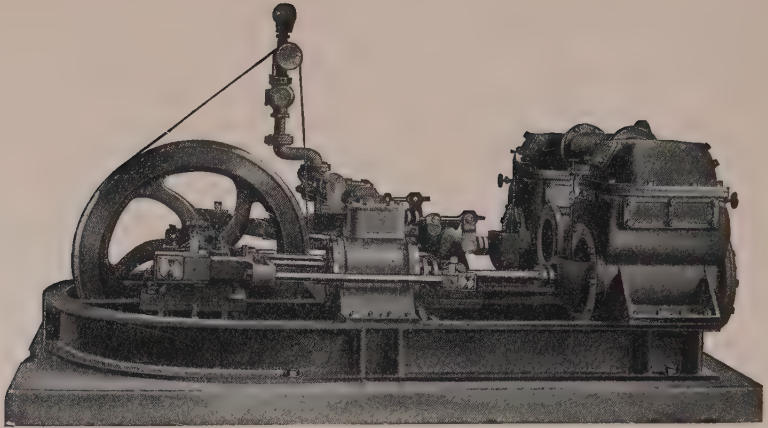
Except where condensers are of the ejector type, as discussed in the chapter on Condensers, they have to be provided with vacuum pumps, whose function it is continuously to remove air and non-condensable gas as fast as these accumulate. In order to do this work, the pump must compress this rarefied air or gas up to the atmospheric pressure, and thus it may be said that a vacuum pump is merely a low pressure air compressor.

The air removed will in general have a large proportion of water vapor mixed with it, corresponding to the temperature as has been pointed out in the above reference.

The range of compression will be from the absolute pressure corresponding to the vacuum maintained up to the atmosphere, or in the average case, from 4 in. absolute (corresponding to 26 in. vacuum) to 30 in. absolute (corresponding to the atmosphere), which is a ratio of $7\frac{1}{2}$ to 1. It may therefore be expected that the re-expansion of air in the clearance of the pump will assume prime importance in its operation. This problem has been attacked from different directions according to the design.

The original pumps for this work were of the "walking beam" type with fly wheels. Afterwards, direct acting pumps were used. These pumps were always operated "wet." In other words, the condensers were installed low down, the injection water being drawn in by vacuum, and both the water and the air went to the pump where they were discharged to the atmosphere mixed together. The design was little different from that of ordinary water pumps, except that the valve area was much larger, and the valves were of soft rubber, with low tension springs. Usually, all stuffing boxes were equipped with "basin heads" for water sealing. (See Fig. 1.)

The clearance spaces in operation were filled with water, and thus the air displaced. In practice, it was found that the mixture of air and water tended to froth and thus form more or less of an emulsion, especially if the pumps were operated too fast. To overcome this tendency, it was necessary to provide in the design ample room in the water space or clearance included between the suction and discharge valves for the proper separation of air and water after their admission into the pump, and to so arrange the shape that the air bubbles should float directly towards the discharge valves. It was also found that better results were obtainable by arranging the flow in such a way that water tends to enter ahead of air, and leaves the pump in the reverse order. In this way, the operation is more positive, as all the clearances are sealed before the stroke begins,



Courtesy Worthington Pump & Machinery Corp.

FIG. 1.—Duplex Horse Shoe Type Vacuum Pump.



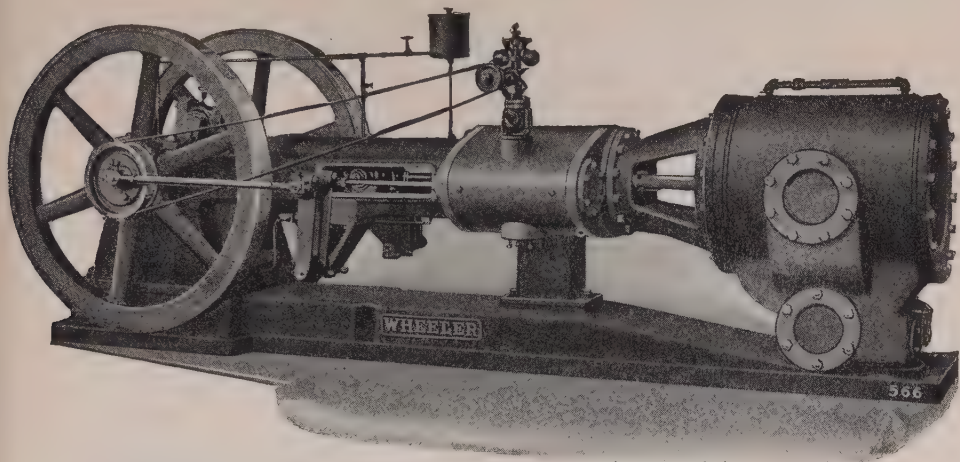
Courtesy Worthington Pump & Machinery Corp.

FIG. 2.—Suction Valveless or Edwards Type Pump.

and are again sealed at the end, thus affording less of a chance for re-expansion and churning.

Direct acting pumps naturally have the obvious disadvantage of a variable stroke, which is very undesirable for vacuum work, and this has led to their abandonment in favor of the fly wheel pump.

The Edwards vacuum pump shown in Fig. 2 was a big step forward in efficiency. Here, there are no suction valves, as these are replaced by the use of large ports around the bottom of the cylinder, which are uncovered by the piston as it reaches the bottom of the stroke. The water and air therefore have free entrance through unrestricted openings, and thus quickly fill the cylinder which is in a vertical position. The discharge valves are in the top head, and let out the air first and then the



Courtesy Wheeler Condenser & Engineering Co.

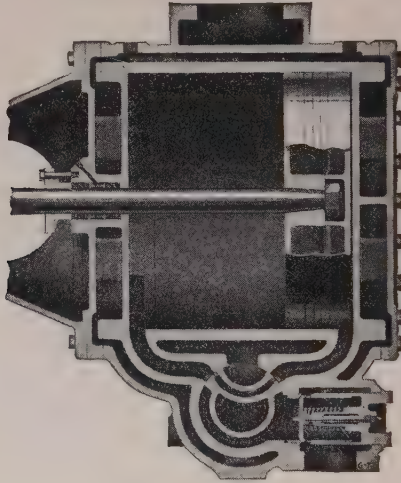
FIG. 3.—Outside View R. D. V. Pump.

water, as suggested above. The pump is only operative on one of its two strokes, the down stroke being used to raise vacuum to a point much higher than that maintained in the condenser. It can operate "dry," namely, for the removal of air only, but in this event, sufficient water must be fed in to seal the valves and clearances. This design has seen many applications, particularly for marine service for which it was particularly designed. It also makes an excellent "sweet water pump" for the removal of condensate from steam belts of evaporators. There have been several modifications of the Edwards idea, some of them of horizontal design, and double acting.

All vacuum pumps in which there must be displacement water must necessarily be operated slowly, and as a result must be of large size for even moderate displacement. In view of the demand for pumps of large capacities and moderate prices intensive study has been devoted to the subject, and the rotative dry vacuum pump (R.D.V.) shown in Fig. 3 has

been designed. Here, the original air compressor idea has been revived, and the pump is in reality only a specially designed air compressor. Instead of having the usual inlet and outlet valves, an air valve has been substituted, much on the order of the slide valve of a steam engine, but of very generous proportions. The use of sealing or displacement water has been done away with altogether, and therefore the re-expansion of the air contained in the clearances again becomes a vital problem. Actually, in the present designs, these spaces are maintained at a minimum.

In nearly all the early R.D.V. pumps, the slide valve was provided with the so-called "trick port" or "flash port," an arrangement by which the two ends of the cylinder are equalized towards the end of the stroke,



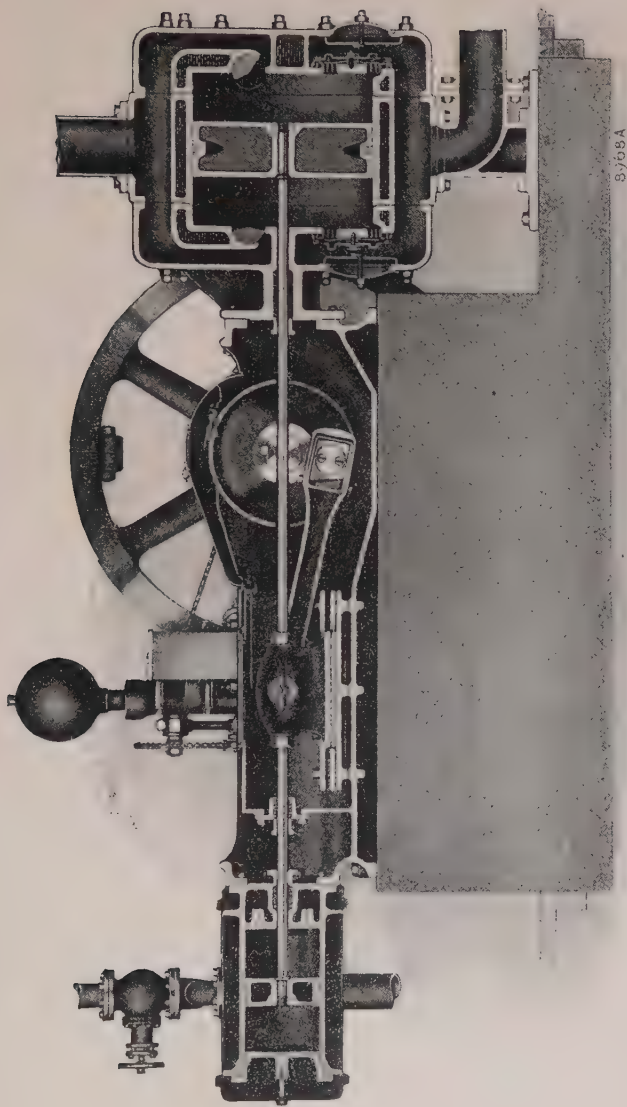
Courtesy Wheeler Condenser & Engineering Co.

FIG. 3a.—Cross Section of Vacuum Cylinder and Valve, R. D. V. Pump.

after the pump is completely closed from the condenser and the atmosphere. In this way, the end of the pump about to begin its new stroke is partially evacuated by the expansion of its contents into the other end before this latter begins its discharge stroke, so that theoretically, at least, the re-expansion of the clearances has been, to say the least, greatly reduced.

The latest R.D.V. pumps, however, have mechanically operated Corliss suction valves, and multiple port featherweight grid discharge valves, and the "trick port" is left out, as shown in Fig. 4. The clearances have been cut down still more, and the operation is better and simpler. The operating speeds are very high, and the first cost has been brought down considerably.

While on the subject of R.D.V. pumps, it is well to again call attention to the variation in the efficiency of displacement of these pumps as



Courtesy Ingersoll-Rand Co.

FIG. 4.—Longitudinal Cross Section through "Imperial" XPV-1 Piston Valve Steam Driven Vacuum Pump

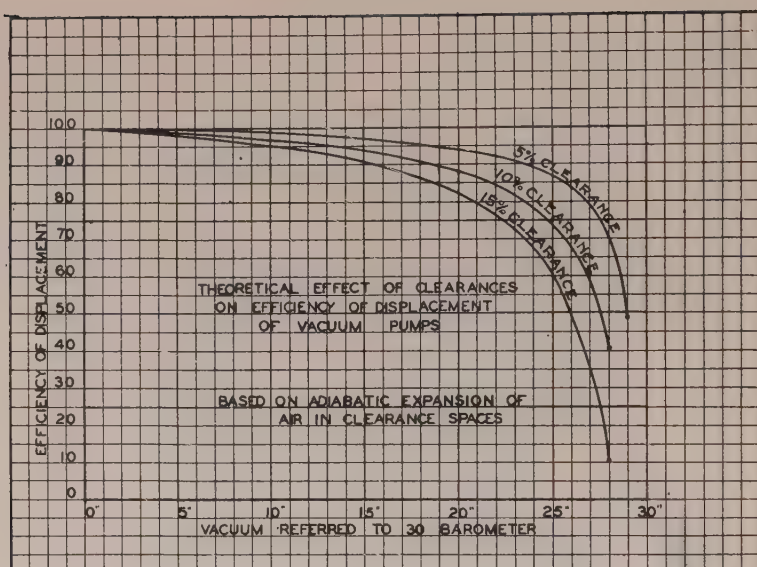


FIG. 5.

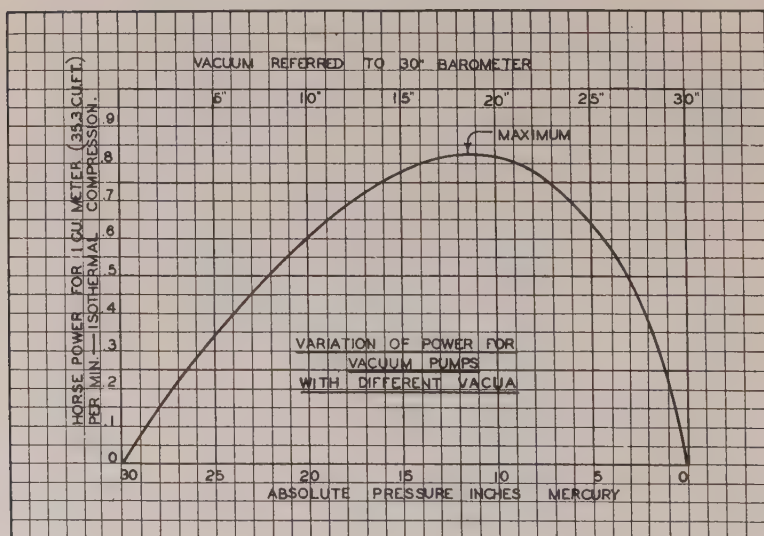
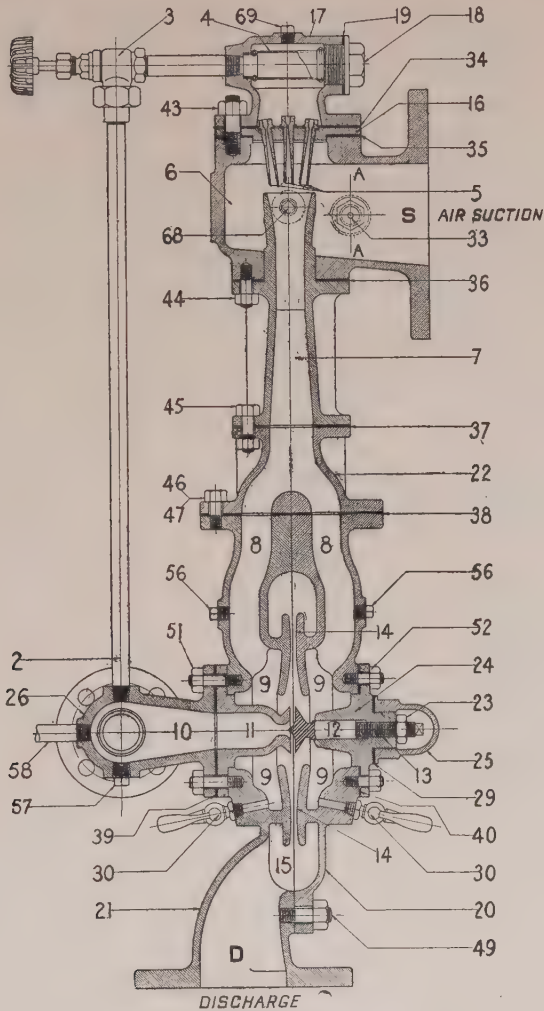


FIG. 6.

compared with the theoretical, as the vacuum maintained is raised. The curve in Fig. 5 shows this characteristic, which must be rigidly taken into account in making calculations for required capacity, as all manufacturers



Courtesy C. H. Wheeler Co.

FIG. 7.—Radojet Air Pump.

rate their pumps on "swept volume," which presupposes 100 per cent. efficiency, as compared with actual performance, which may drop to as low as $33\frac{1}{3}$ per cent. at high vacuum.

It is also well to reiterate here that the maximum power consumption of vacuum pumps is not at highest vacuum, but at some intermediate point in the neighborhood of 18 in. to 20 in. vacuum, as shown in Fig. 6. This will be readily understood by studying an indicator card taken from a vacuum pump at various vacua. The point mentioned above gives the largest mean effective pressure, and therefore requires the greatest amount of power.

Within the last few years, a new type of vacuum pump has entered the field, and has given good results, particularly at high vacua. This is the steam jet pump, shown in Fig. 7, which is nothing more than a carefully designed steam ejector. There are no moving parts, and this unit constitutes a very simple, efficient, and inexpensive piece of apparatus. Live steam is used in the nozzles, and as it expands through the throats, it entrains the air, which becomes mixed with it, the two being discharged to the atmosphere. The only problem in connection with the use of this pump is the utilization of the steam discharged from it. It cannot be returned to the exhaust system without impairing the operation of steam consuming apparatus which would soon become partially "air bound." This mixture of steam and air can be used for heating cold solutions where ample surfaces are provided. One word of caution: this mixture is quite corrosive, and care must be used in providing against this danger, otherwise, pitted tubes will be the result. This is doubly important if the injection water going to the condensers contains corrosive gases, such as hydrogen sulphide.

There are several other types of vacuum pumps which have been used with more or less success, but it is beyond the scope of this book to elaborate further. The reader is referred to works on pumping machinery and on condensers.

In making selections for size ample allowance should be made for leaks in evaporating apparatus, which have a great many joints, valves, stuffing boxes, and fittings. To be safe, the vacuum pump capacity should be about twice the size advisable for turbines operating on the same load and vacuum conditions.

Chapter II.

Evaporator Accessories.

All evaporating apparatus should be amply provided with all the instruments and accessories required, in order that the operation may be carefully checked, not only for the purpose of keeping up the efficiency, but also for actually improving it when the opportunity presents itself. Without proper indications, it is impossible to check anything, and the work becomes a matter of pure guessing. Accordingly, there is given below a list of such accessories with comments where it is deemed appropriate.

1. Gauges. Pressure and vacuum gauges must be provided at all points of interest, including the steam belt of the first effect, and all the vapor belts. The pressure gauge may be of the spring-dial type, but the vacuum gauges must be of the mercury column design. Spring vacuum gauges have a marked tendency to read off, particularly after rough usage, or exposure to heat, as the springs are very weak, and soon yield under these conditions. For accurate information, one should not count on spring vacuum gauges.

Mercury gauges must be protected with a trap to catch the water which condenses in the pipe connections. The "U"-Type is preferable and can be used for either pressure or vacuum within the available range. Such gauges usually have a sliding scale by the use of which the difference of level of the mercury in the two branches of the "U" is actually measured, which gives an accurate reading, providing no water has been allowed to accumulate on top of the mercury.

2. Thermometers. Thermometers should be provided immersed in the boiling liquid in each body of the multiple effect. They should be calibrated to read temperatures on one side of the scale, and the corresponding steam pressures and vacua on the other side. This constitutes a very good check on the vacuum gauge, unless there is an abnormal rise in the boiling point. This, however, would only cause a discrepancy of about 1 in. at the outside in the average case, and that only on the last effect. There are exceptions, such as caustic soda, which has a very high temperature rise, calcium chloride and others.

There should also be thermometers in the vent pipe of each steam belt, on the evaporator side of the control valve, for indicating the partial vapor pressure, thus showing whether the vent is sufficient.

3. Level Gauges. Water level gauges must be installed in each steam belt to indicate whether or not the condensed steam is being removed.

A well-designed water column must be provided spanning the steam belt of each effect, indicating the level of the liquor in the evaporator.

The words *well designed* cannot be exaggerated. Most of the fixtures purchasable from stock are useless under vacuum, the result being that air bubbles enter all the little stuffing boxes, and flow up the glass tube, thus so disturbing the apparent level that it becomes impossible to make a reading. So these fixtures must be carefully selected, and made up "bottletight," otherwise, they are of no service.

Incidentally, it is well to connect the upper part of the column considerably above the probable liquor level in the evaporator so as to avoid disturbances. In putting in the rubber gaskets on the glass tube to fit into the stuffing boxes, turn this tubular gasket inside out before slipping it on the tube. It will be found that on account of the curved shape it assumes, it is much easier to make tight.

4. Density Testers. There must be provided in the last body of each multiple effect a density tester by means of which the concentration of the liquor is determined. This can be made very simply by having a 2-in. pipe connection just above the upper tube sheet, coming out horizontally, with a gate valve. This connects into the side of a little vertical chamber about 4 in. diameter and 12 in. high, near the top. This chamber has a petcock on the top, and a $\frac{1}{2}$ in. connection with valve at the bottom. By opening the 2-in. valve, closing the petcock and the $\frac{1}{2}$ -in. valve, the chamber quickly fills with liquor from the evaporator, when the 2-in. valve is closed, the petcock opened, and the liquor drawn off through the $\frac{1}{2}$ -in. connection at the bottom into a test cup for proofing, after which it can be returned to the evaporator, if desired, by a permanent connection to the chamber referred to above.

5. Sight Glasses. There should be a generous number of sight glasses in each body of the evaporator to observe its operation on the inside. These glasses must be so designed that they come about flush with the inside of the body if possible. They should be made of heat-resisting glass and at least $\frac{1}{2}$ -in. thick. Ordinary plate glass breaks very easily with the violent temperature changes. The "bulls eyes" must not be too small, at least 6 in. in diameter in the clear, and preferably larger. The edges of the glasses where the joint is to be made must be ground to a uniform thickness, otherwise there will be breaks when the joints are made. When installing, thick soft gum gaskets must be used on both sides of the glass to avoid strains.

6. Vacuum Breaker. Each effect must be supplied with a vacuum breaker, which is merely a valved connection to the atmosphere. Many have tried to make these devices noiseless in operation, which is practically an impossibility, and not worth the effort it entails.

7. Manheads. There must be a manhead just above the top tube sheet and in the bottom of each body of all multiple effects. These must be large enough, from 16 in. to 18 in. in diameter. They must have hinges with yokes and handwheels for tightening, and the whole design must be studied for ease and quickness of operation, and freedom from trouble of any kind. Slotted holes must be put in the hinges, otherwise the wearing of the gasket will make it impossible to tighten under the hinge.

8. Safety Valves. As a matter of precaution, each steam belt should

have a small pop safety valve located in the rear, preferably of brass, and of the sealed spring type, set slightly above the operating pressure. This is in addition to the master safety valve usually supplied on all exhaust systems.

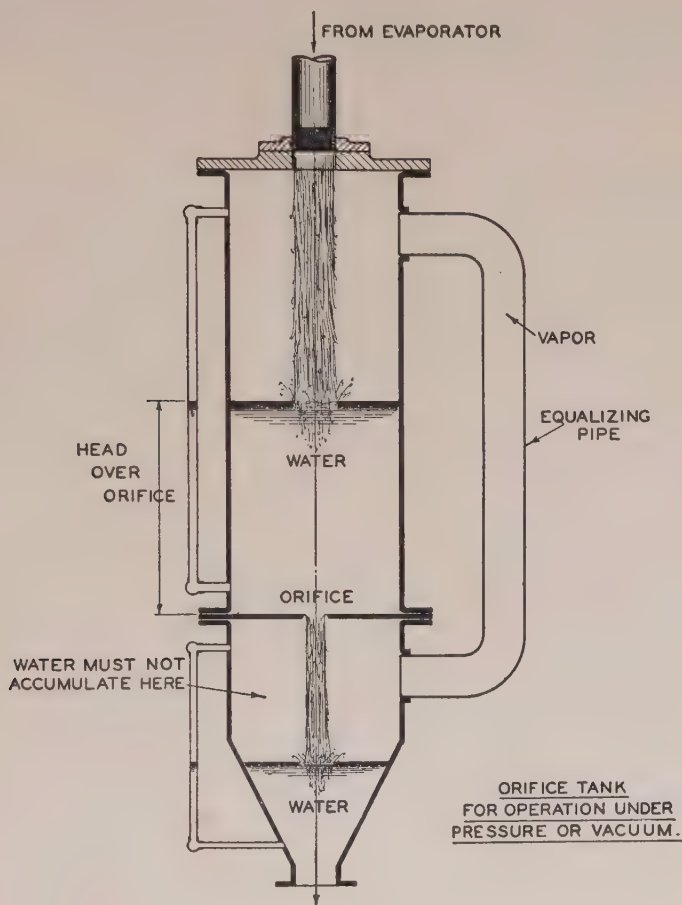


FIG. 1.

9. Steam Meter. A very useful, although not indispensable adjunct to any evaporator is an indicating orifice tank showing the quantity of condensed steam leaving the first effect, which is in reality a direct measurement of the speed of the evaporator. Usually this can be made very simply on the job from scrap pipe and fittings. The flow of water through orifices is given in all handbooks for different heads. Fig. 1 shows such an orifice tank. The orifice must be calibrated for precision.

10. **Pressure Control.** Evaporators operate best when fluctuations are eliminated. It is therefore advisable to use properly designed and proportioned pressure reducing valves between the live and exhaust steam lines in order to maintain a constant pressure on the exhaust system. It is best to have a by-pass to the valves, and not make them too big. They never operate properly unless they are well loaded, and many such a control has failed because of neglect or ignorance of this factor.

Chapter 12.

Insulation.

The main difference between the actual steam consumption of a given evaporator and its estimated steam consumption, as shown by a heat balance calculated according to the actual operating conditions, is due to heat losses to the surrounding atmosphere.

The exposed surfaces are considerable, and therefore one may expect serious loss from this source. The greatest loss occurs in the first effect, for not only will the radiation be greater on account of the higher prevailing temperature, but such a loss will be of more consequence, inasmuch as the heat so dissipated has only been used for one stage of evaporation.

There have been many tests made to establish the facts, and some of the data is contradictory. The losses will be more serious as the amount of radiating surface is great as compared with the capacity of the equipment. Thus the greater the number of bodies, the larger the proportional loss. Also, for a given evaporator, the losses being practically constant, the slower its operation, the greater the proportional loss of heat.

Accurate tests were made on a 9000 sq. ft. long tube quadruple effect having a very fast rating, and without insulation. The performance was 95 per cent. of what it should have been as indicated by a carefully calculated heat balance. It is believed, however, that in this particular instance, the losses were much lower than the average, which would be about twice the figure.

E. W. Kerr, after studying the subject from many angles and after many tests in the sugar industry, estimates the loss of heat in multiple effects in percentage of steam delivered to the first body as follows:

Evaporator	Uncovered	Wholly Covered	Partly Covered
Double	1.06	0.26	0.46
Triple	4.20	1.05	2.07
Quadruple	9.80	2.70	5.00

It is believed that the losses in practice for uncovered evaporators may run even higher, although these figures represent what might be considered a fair average. It is therefore evident that it pays to insulate.

In particularly warm climates, the heat around a large uncovered evaporator is unbearable, and for the comfort of the operators alone, insulation should be applied.

It is advisable not to put on the covering until the equipment has been

thoroughly tested with steam, and all the leaks taken up. Otherwise, it may be necessary to rip out covering to tighten defective joints.

As to the kind of insulation, this need not be of the most expensive type, as the temperatures are relatively low; 1½-inch asbestos air cell is very good. This can be covered over with wood lagging, thus protecting against injury, and maintaining a tidy appearance.

It may not pay to cover up completely. Some make it a practice to insulate the large parts only, particularly where radiation would tend to annoy the operators.

Chapter 13.

Investigating Troubles.

Evaporators, in common with equipment of all kinds, have their troubles, and the most difficult task is to investigate and rectify their deficiencies. Naturally, success in this work depends to a great extent upon experience and mental aptitude.

Open Mind. A few broad principles must be observed, the foremost of which is to keep an open mind at all times, studying the problem from different angles, so as to obtain several points of view, eliminating possibilities only when justified by evidence against them, rather than arriving at conclusions not substantiated by indisputable facts. This is very important to any one carrying on research work; and in spite of all recognized procedure to the contrary, a large amount of time, energy, and money is spent constantly, following out ill-matured conceptions founded on pet ideas, only to find out ultimately that the real trouble was elementary, and could have been rectified easily, if the investigator had looked for it, instead of making up his mind beforehand.

Keen Observation. Another most valuable asset is a keen sense of observation. Just as a patient usually exhibits symptoms of his disease, so does an apparatus show signs of the cause of its failure; and a successful diagnosis depends upon observation and correct interpretation of such evidence.

It is usually wise first to make a rough general survey of the situation, paying particular attention to anything that appears out of the ordinary, and intensifying on such points if they are apparent. Failing in this, the next step is to follow a rigid process of elimination, starting with the easiest possibility, and passing on, as they are discarded in turn.

Fundamental Essentials. There are really but few fundamental conditions necessary for the proper operation of evaporators, and difficulties arise from neglect of these. Perhaps it would be well to elaborate.

Foul Surfaces. In order to secure good transmission of heat, it is necessary that the heating surfaces be kept reasonably clean. It is suggested, therefore, that the condition of the heating tubes be examined first. These invariably foul more at the bottom than at the top, and so, when the observation is made, it must not be taken for granted that because the upper ends are clean, that this cleanliness obtains throughout. By putting an electric light in the bottom of the evaporator, and looking through from the top, a very fair idea of the condition of the surfaces may be obtained. The fouling of tubes, the bad effects, the cause, and the usual remedies have been discussed at length in another chapter.

Feed Pipes. The next simplest step is to make sure that all feed pipes, condensate pipes, and vent pipes are clear. It is well to bear in mind that the liquor velocities in such pipes must not be too great. In ordinary multiple effects, the upper limit is about five feet per second. If the quantity of liquid passing through imposes a speed greater than this, difficulty will be experienced, until the said pipes are enlarged to reduce this velocity.

Water Removal. It is, of course, absolutely necessary to remove continuously and positively all of the water of condensation from the steam compartments or calandrias; if not, there is a progressive accumulation which will eventually fill the steam side of the tubes, thereby blanking off the equipment resulting in a decrease of capacity, and in extreme cases, an ultimate stoppage.

Many evaporators are in operation to-day in which no provision is made to indicate whether or not this condition exists, and it is merely a matter of guess work in this determination. Water gauge glasses must be installed in each calandria which indicate the level of water above the bottom tube sheet.

Air Binding. Similarly, it is indispensable that air and non-condensable gases be removed from the steam belts, and if the apparatus is not so equipped, it should be provided with vent pipes, properly located in the top tube sheet, and connected to a header outside discharging into the vapor space of the last body.

If individual vent pipes extend too close to the bottom tube sheet, it is easily possible for these to pick up water, discharging it into the last body, thus diluting the concentrated liquor, and this detail must therefore be carefully looked into.

Air Pockets. The location of the individual vents must be such as to completely remove the non-condensable gas in the calandrias, otherwise, that portion of the heating surface where pockets or zones of gas accumulate will become inactive, with a corresponding loss of capacity. Such defects can usually be determined through the sight glasses while the apparatus is in operation, and remedied accordingly. These spots can be located by lowering the boiling levels, when, due to the mechanical spouting of the active tubes, they can easily be distinguished from the inactive ones. Generally speaking, the gases to be removed are heavier than steam, and will therefore be found near the bottom. This is not universally true, however, and it very often happens, especially if the solution to be evaporated is of organic origin, that ammonia is liberated, which, being lighter than steam, will accumulate just under the top tube sheet.

Air Leaks. Perhaps the most common source of evaporator troubles is loss of vacuum through air leaks. The easiest and most positive way to locate them is by filling the bodies with water, when any opening of consequence will be revealed at once.

The presence, but not the location of leaks, may be readily established by observation, for they are always accompanied by a corresponding loss of vacuum, due to the inability of the vacuum pump to remove the increased volume, thereby partially air binding the condenser. In a multiple effect,

the leaks may be in any one of the bodies which normally operate under vacuum. It is easy to locate the particular leaky cell as follows:

Close all the pipes for feed, drains, vents, washout, manheads, etc., and raise vacuum. If the usual 27 inches is not obtainable, the leak is in the last body.

If fair tightness is established here, next open the vent to the steam belt of this calandria. This will raise vacuum not only at this point, but through the vapor pipe, it will also evacuate the liquor space of the preceding body. Thus it can be established whether the leak is at this point. Similarly, the same procedure can be used for the preceding cell, and so on, until the defect is finally located.

Sometimes, due to porous castings, it is not possible to obtain the necessary vacuum, even though all methods of locating leaks have been tried. If such is the case, it will be noted that the vacuum pump is discharging a considerable volume. The remedy is a coat of suitable paint applied under vacuum.

Location By Sound. It is well to call attention to a very interesting combination having to do with the squeaky sound made by a vacuum leak. Under certain conditions of size and shape of opening, it is not only possible, but probable that such leaks will be almost noiseless at vacua exceeding a certain critical point, at which the speed of air entering the orifice is greater than the speed of sound, which we know to be about 1080 ft. per second under standard atmospheric conditions. Therefore, when endeavoring to locate vacuum leaks by their sounds, it is better not to go too high with the vacuum. This is doubly true if the evaporator is in a plant where there is considerable machinery in motion, such as turbines, motors, engines, elevators, or conveyors.

Air Leaks and Pockets in Pipes. Another source of trouble usually met in evaporators is the removal of condensate from steam belts under vacuum, and also the removal of concentrated liquor from the last body, which is virtually the same problem in an aggravated form. This has been discussed in another chapter, Condensate Removal, but there are certain principles involved which it is well to bring out here.

The statement has many times been made that one of the most elusive phenomena encountered in this work is the behavior of the flow of hot or cold liquids through pipes where such liquids are mixed with vapor or non-condensable gas, especially under partial vacuum, all of which conditions obtain in both the feed and condensate pipes of an evaporator.

One all important detail must be carefully looked after: namely, no such pipe or pipe line should have pockets, or high points between two lower points on either side, for such a combination invariably forms a trap where air or vapor lodges interfering with the free passage of the liquid. If due to local conditions the formation of a pocket of this kind is inevitable, then the upper part of the pipe should be tapped, and vented back to the source.

Air leaks in condensate and feed pipes of evaporators will cause untold troubles, for the volume of this air is swelled enormously, not only by the lowered pressure, but by the partial vapor pressure of the hot liquid or

condensate, thus blocking the system, and disrupting the flow. It is wise, therefore, to make sure that there are no such leaks.

Water Hammer Feed Pipes. The flow of relatively cool liquid through a pipe system only partially filled, with accompanying liquor vapor, is likely to cause very serious trouble from water hammers. We have had several experiences with overflow level controls in multiple effects, where the control chamber was outside of the body. Such disturbances are likely if the equalizing pipes are too small, particularly when starting up, before the entire contents of the body have been thoroughly heated, or if the entering feed is of a relatively low temperature. In the overflow outside control, this difficulty is caused by the annular sheet of liquid dropping down in an atmosphere of steam, the steam being hotter than the liquid, when condensation takes place faster than the said steam can be supplied through the equalizing pipe. The remedy is a larger equalizer.

Leaky Tube Sheets. The tubes and tube sheets of an evaporator should be water tested at regular intervals for leaks. Such a leak in the upper tube sheet is not of much consequence, inasmuch as the net result is merely a shunting of a small amount of steam. In the lower tube sheet, however, this becomes a matter of prime importance, for the lower part of the calandria, being covered with water of condensation, the condensate coming through dilutes the concentrated liquor, thereby destroying work done by the apparatus, requiring that much additional evaporation over and above the ordinary requirements. This is emphasized if the difference of pressure between the two sides of the heating surface is great.

Barometric Seal. If the apparatus is provided with a barometric condenser, one must be sure to fill the hot well, and also to keep the end of the pipe sealed at all times, otherwise, the vacuum will be broken at this point.

While on the subject, it is well to call attention to the necessity of keeping the barometric column as direct as possible, preferably without any horizontal runs, for it must be remembered that a considerable quantity of air is entrained through this pipe, and that precautions against air pockets referred to above is doubly important. We have seen an air bound tail pipe to a barometric condenser, where this pipe dropped down vertically to the ground, and then went horizontally about 150 feet to a central hot well. The injection water backed up into the condenser to a point high enough to materially interfere with the vacuum, producing a net loss of 2 inches.

Forced March. There is a most curious misconception on the part of some multiple effect operators in the sugar industry. In this case, where due to fouling, the evaporator cannot concentrate to the proper density all the juice of the factory, these men resort to what is called a "forced march," which consists of carrying high levels, thus storing momentarily a certain proportion of the juice in transit, but actually slowing up the apparatus very considerably. (There has been discussed thoroughly in other pages the effect of high levels.)

It is almost impossible to make these operators realize the folly of their methods, and the fact that they are actually making a bad condition much worse, and blocking themselves hopelessly. The only way to run an evaporator properly is to carry the level at such a point as to just cover the tube sheet, unless the equipment is used for a crystallizing solution, in which the conditions change radically. This is discussed separately.

SECTION III.

Chapter I.

General Applications to Specific Industries.

In the third section of this book, it has been attempted to give useful information as to the intelligent application of the principles of multiple effect evaporation to various related industries.

More space has been devoted to the sugar industry than to the others in view of the large number of complex problems existing and their great industrial importance. The methods and processes brought out will no doubt be very illuminating to other industries in which the developments of multiple effect evaporation have not been carried out with so much thoroughness.

A number of other problems have been covered in a general way, and it is the intention of the author in case there is a sufficient demand to justify a later edition of this book, to further elaborate and amplify the subject.

Chapter 2.

The Heat Balance in the Cane Sugar Factory.

The problem is to make a balance between the available heat derived from the combustion of bagasse and other fuels, and the heat consumed by the plant, which, in general, is limited to five main requirements as follows:

1. **Power.** Enough power must be generated to operate the mills, pumps, centrifugals, and other accessories, and thus is used the corresponding heat equivalent of this power.

2. **Heating.** All of the cold juices entering the factory must be heated from their initial temperatures to approximately the boiling point in the process of defecation.

3. **Evaporation.** Approximately 75 per cent. of the hot juice must be evaporated after defecation to convert it into syrup for use in the vacuum pans. Heat is here used in multiple effect.

4. **Crystallization.** The syrup, after leaving the evaporators, goes to the vacuum pans where it is converted into sugar and molasses by the use of steam in single effect. In this station it is necessary to evaporate not only most of the water left in the syrup but, in addition, the water added to the molasses for dilution in working down seconds and thirds.

5. **Radiation.** A considerable portion of our available heat supply is lost by radiation, the amount depending upon the design of the various steam and exhaust pipes, hot juice and syrup containers and the manner in which they are insulated against heat losses.

Concrete Example. We will assume for consideration and analysis the specific case of a sugar factory in which 100,000 lbs., or 4,000 arrobas of cane are ground per hour. This is perhaps a small unit but the calculations can be easily altered for any other size by multiplying by a simple factor. We will also assume that with normal maceration the weight of the dilute juice will be equal to the weight of the cane ground; in this case, as mentioned above, 100,000 lbs. per hour. Warm sweet water is used on the mills and the initial temperature of the juice will therefore be taken at 100° F., which is fairly representative. We will assume further that the conditions of exhaust pressure and vacuum will be standard, namely—5 lbs., and 26 in., referred to a 30-in. barometer.

Heat Equivalent of Power. The actual power requirements of a sugar factory vary considerably with the design but we will assume 1,500 horsepower as a fair average for a factory of the capacity set forth above. 2546 B.t.u.'s are equivalent to one horsepower hour. Therefore, the heat equivalent of 1500 horsepower hours is 3,820,000 B.t.u.'s, which corresponds to the condensation of 3940 lbs. of steam from and at 212° F.

This represents the heat equivalent of the power, indeed, a very small fraction of the total required.

The steam which has passed through the power generators, where it has given up this 3,820,000 B.t.u.'s, is available for use as exhaust steam and contains the total original amount of heat less the small quantity used for power. It is, of course, further depreciated by radiation losses for which allowance must be made and its temperature has been reduced from that corresponding to the live steam pressure down to 227° F. the exhaust temperature.

Juice Heating. The heating of raw juice consumes a very much larger quantity of steam than would be supposed without careful calculation. It is necessary to raise the temperature of 100,000 lbs. of juice per hour from 100° F. to 212° F. This requires 11,200,000 B.t.u.'s, an amount of heat which corresponds to the condensation of 11,540 lbs. of steam from and at 212° F. per hour. Heating is done with exhaust steam at 5 lbs. pressure or with vapor and exhaust or in some cases with live steam. This last, however, is exceptional in modern practice.

In making the above calculations, we have assumed the specific heat of juice as 1.00. As a matter of fact it is slightly less than 1.00 but the intentional slight error simplifies not only these calculations but subsequent ones as well and introduces a slight margin of safety which is not undesirable.

Evaporation. At the multiple effects 75,000 lbs. of water must be evaporated from 100,000 lbs. of juice, thereby leaving 25,000 lbs. of syrup ready for the pans. Due to radiation losses during the settling period the temperature of this defecated juice will be less than its temperature just after heating. Under ordinary conditions, such as we will assume here, we may expect this to be about 180° F., having dropped from 212° F., a loss of 32° F. With careful insulation against radiation, such as is provided with the *Dorr Clarifiers*, the loss will not exceed 6° or 7° F. and the defecated juice temperature will be approximately 205° F. The steam consumption of the evaporators is very much affected thereby, as is clearly shown on the performance diagrams in the following pages.

The full discussion of the various cycles used will be given later.

Vacuum Pans. Opinions and actual records of the amount of heat required by the vacuum pan station vary considerably. In a well-operated factory, however, it may be assumed that the amount of steam necessary for the work contemplated here should not exceed 20,000 lbs. per hour from and at 212° F. After properly designed and proportioned equipment has been installed, economy is strictly in the sugarmaker's hands. The only advice we can give is to hold the dilution of molasses down to a minimum and to use hot water for this purpose.

The lag, or interval of time between the evaporators and vacuum pans, must be maintained as short as possible and heat must be conserved by properly insulating hot syrup containers wherever possible.

Radiation. A most important and elementary fact to be carefully borne in mind about sugar factory operation is that *the supply of heat is limited* and unless the steam consumption can be maintained within the

amount available from the combustion of bagasse alone, this inefficiency necessitates the burning of expensive additional fuel.

All sources of heat losses must be carefully checked. Not only should all live and exhaust steam pipes be covered but also all containers, tanks, pans, and power auxiliaries, where there may be any heat losses. It is only by rigid conservation that the best results are obtained.

If carefully controlled, the loss of heat in the factory under consideration, should normally be less than 2500 lbs. of steam from and at 212° F. per hour.

Improvements in the Use of Fuel

The present practice in sugar factories is the result of many years of development, starting from the old direct-fired open kettles, in which only artificial fuel was used. Originally, the bagasse was disposed of by burning in a large refuse burner so that the heat from its combustion was entirely wasted. Indeed, it was customary in the old days to help dispose of the bagasse by burning wood with it, so as to maintain combustion. Those were the days of the three-roller mills and the small factories. As space does not permit our making an historical review of the developments, suffice it to say that the changes were brought about gradually, so that to-day, instead of using additional fuel to dispose of the bagasse, this fuel alone, under favorable conditions, generates enough steam to run the factory. This has been made possible by reducing the moisture content of bagasse by better extraction, also by the development of furnaces adapted to the work required. In addition, multiple effect evaporation has been improved to a point where the heat requirements of the factory have been reduced to within the range of the available steam supply. Present day conditions are fairly well represented in the following example:

Heat Value of Dry Bagasse. The consensus of opinion as a result of many tests has shown that the heat of combustion of dry bagasse is about 8300 B.t.u.'s per pound. We have used this figure in making up the calculations which follow for bagasse boiler performance.

When bagasse is fired wet, as is always the case, the heat available per pound as fired is proportional to the weight of the dry contents. The water occurring as moisture must be heated to 212° F., evaporated into steam at that point and then superheated, leaving the boiler at the flue gas temperature, assumed here at 500° F. In our calculations the heat expense of these physical changes must be deducted from the total heat available.

Assuming that bagasse is admitted to the furnace at 100° F., for each pound of moisture contained, the heat expense will be approximately as follows:

(1) To heat 1 lb. water from 100° to 212° F.....	112.0 B.t.u.
(2) To evaporate 1 lb. water from and at 212° F.....	970.4 B.t.u.
(3) To superheat same from 212° to 500°, 288° F. 288×0.463 (spec. heat suph. steam).....	133.3 B.t.u.

Total to dispose of 1 lb. bagasse moisture.....	1215.7 B.t.u.
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We submit, herewith, a log in which probable conditions have been taken into account and a graphical chart from which it is possible to interpolate for any intermediate point.

TABLE SHOWING RESULT OF BURNING ONE POUND BAGASSE UNDER VARIOUS CONDITIONS.

Moisture content	42%	44%	46%	48%	50%
Dry content	58%	56%	54%	52%	50%
Heat from combustion B.t.u.'s.....	4820.	4650.	4480.	4320.	4150.
Heat to evaporate water B.t.u.'s.....	511.	535.	559.	583.	607.
Heat available for use B.t.u.'s.....	4309.	4115.	3921.	3737.	3543.

Equivalent evaporation

From and at 212° F. lbs.....	4.44	4.24	4.04	3.85	3.65
Eq. steam 70% eff. boiler lbs.....	3.10	2.97	2.83	2.70	2.55
Eq. steam 65% eff. boiler lbs.....	2.88	2.76	2.63	2.50	2.37
Eq. steam 60% eff. boiler lbs.....	2.66	2.54	2.43	2.31	2.19
Eq. steam 55% eff. boiler lbs.....	2.44	2.33	2.23	2.12	2.01
Eq. steam 50% eff. boiler lbs.....	2.22	2.12	2.02	1.92	1.82

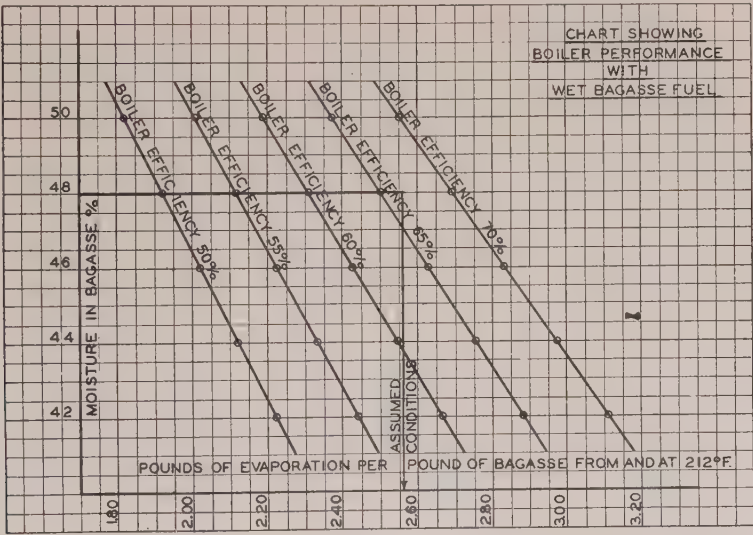


FIG. I.

The conditions which we have assumed for the specific case, as indicated in our chart are actual conditions of one of the well-operated Centrals in Cuba. The evaporation from a pound of bagasse is 2.56 lbs. from and at 212° F. and the bagasse contains 48 per cent. moisture, indicating a boiler efficiency of 66⅔ per cent.

Steam Supply. Except where the fibre content of the cane is very low, *i.e.*, less than 10 per cent., the bagasse after leaving a modern mill should provide sufficient fuel for the operation of the sugar factory. It is generally true that the fibre content of cane grown in new soil is low, necessitating the burning of a certain proportion of additional fuel, particularly when liquidating or when the cane supply to the mill is irregular and stoppages are frequent.

As mentioned previously, average fair conditions are represented by a fibre content of 10 per cent. The factory under consideration, therefore, would have 10,000 lbs. of dry fibre plus solids not removed by the mills. These solids consist of sucrose, glucose and other organic and inorganic matter amounting to about 1000 lbs. The total weight of dry bagasse would be 11,000 lbs., which, however, would not go to the furnaces as such but, on the contrary, would contain probably 52 per cent. dry material and 48 per cent. moisture, making the total weight of bagasse 21,100 lbs. per hour.

The amount of steam which can be generated by burning this fuel depends not only upon the design of the furnaces but also upon the care and attention with which such furnaces are kept up and operated. Under favorable conditions, 1 lb. of wet bagasse as specified above gives 2.56 lbs. of steam from and at 212° F., showing a boiler efficiency of 66 $\frac{2}{3}$ per cent. As the total fuel per hour is 21,100 lbs., the steam available from bagasse alone should be about 54,000 lbs. per hour, which probably represents a maximum figure. Naturally, a change in the fibre content of the cane or the moisture content of bagasse will be followed by a corresponding change in the available steam supply.

Perhaps the greatest factor militating against efficient operation of the bagasse furnace here is the admission of an excessive amount of cold air either through the bagasse chute or through the ash pit doors. In order to obtain good results it is also very necessary to regulate the supply of fuel so that each boiler will take care of its share of the load.

It has been established by numerous experiments and tests that, for efficient operation, bagasse must be burned alone. It must not be mixed in the furnace with other fuels, such as oil, coal or wood, a practice often followed. In cases where there is an insufficient steam supply and additional fuel must be used, much better results are obtained by burning such fuel under separate boilers with settings designed for the particular fuel used and with draft conditions adjusted accordingly. As it is not the intention of this discussion to go into details on the subject the above comments are offered as incidental suggestions only.

Various Cycles Used

There are various cycles in use to-day in cane sugar factories in connection with the evaporating and heating operations which will now be considered in detail. As a matter of convenience in discussion and reference, these cycles will be identified as follows:

- A { Heating with live or exhaust steam;
Evaporation in Triple Effect;
Vacuum pans with live or exhaust steam.
- B { Heating with vapors from first effect, and exhaust;
Evaporation in Triple Effect;
Vacuum pans with either live or exhaust steam.
- C { Heating with live or exhaust steam;
Evaporation in Quadruple Effect;
Vacuum pans with either live or exhaust steam.
- D { Heating with vapors from first effect, and exhaust;
Evaporation in Quadruple Effect;
Vacuum pans with either live or exhaust steam.
- E { Heating with live or exhaust steam;
Evaporation in Quintuple Effect;
Vacuum pans with either live or exhaust steam.
- F { Heating with vapors from first effect;
Evaporation in Quintuple Effect;
Vacuum pans with either live or exhaust steam.
- G { Heating with live or exhaust steam;
Evaporation in Quadruple with Pauly;
Vacuum pans with live or exhaust steam.

The analyses of the various cycles given in detail are based on the capacity outlined above, namely, a grinding rate of 100,000 lbs. cane per hour producing 100,000 lbs. of dilute juice from which 75,000 lbs. are evaporated in the multiple effect.

The intermediate temperatures and pressures in the evaporators have been assumed to represent fair practice when operating under the probable conditions of 5 lbs. pressure in the steam belt of the first body and 26 in. vacuum in the condenser.

Cycle "A." Cycle "A," in which evaporation is done in triple effect and juice heated with live or exhaust steam, was at one time in very wide use in Cuba and many other cane sugar countries. It is very simple in operation and many factories continue to use this method.

The estimated minimum steam consumption per hour is as follows:

	Pounds of Steam F. & A. 212° F.
(1) Power	3,940
(2) Heating	11,540
(3) Evaporation	26,500
(4) Vacuum pans	20,000
(5) Radiation	2,500
(6) Total	64,480
Available from bagasse alone (10 per cent. fibre).....	54,000
Deficit to be made up with other fuel.....	10,480

If the heat in the juice is conserved and the feed to the evaporator is 205°F. instead of 180°F. , as estimated, the reduction in the steam consumption of the evaporator is 2570 lbs. per hour. The deficit is then 7910 lbs. from and at 212° (see chart accompanying).

The heat balance submitted herewith has been calculated on the basis of juice entering the equipment at 180°F. However, in order to estab-

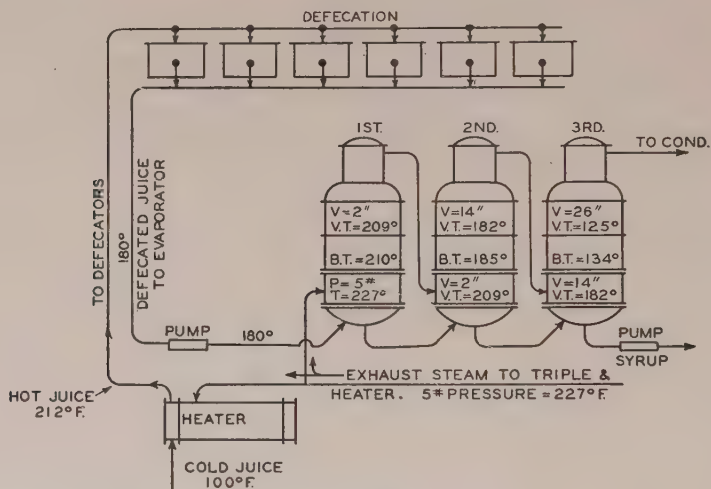


FIG. 2.—Cycle "A"—Triple Effect with Exhaust Heater.

lish the chart which shows the performance with reference to this variable, it is necessary to secure at least two additional points, one for 140°F. and one for 220°F.

The only variation imposed by these changes is to increase or decrease the steam admitted to the steam belt in order to raise the juice to the boiling point as it exists in the first effect.

This means, in the first case, the juice must be heated 40° more than the given heat balance indicates. This involves the expenditure of $100,000 \times 40 = 4,000,000$ B.t.u.'s, equivalent to condensing $\frac{4,000,000}{970.4} = 4120$ lbs. of steam from and at 212°F.

In the case of feed at 220°F. the correction is obviously negative and equal in quantity, inasmuch as the temperature change is the same in amount, *i.e.*, 40°F. but in the opposite direction.

The relative evaporation in the different bodies remains the same and the curves are, therefore, plotted accordingly.

The same corrections apply to all other cycles discussed except cycle "G," which is treated separately.

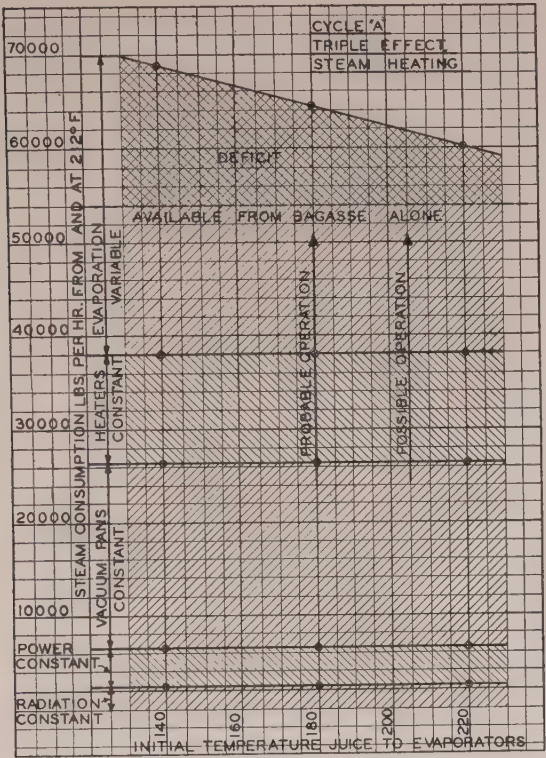


FIG. 3.

TRIPLE EFFECT HEAT BALANCE—CYCLE “A.”

Bodies	Specifications	Heat B.t.u.'s	Juice, Pounds
(1) Steam, 26,800 lbs. @ 227°	$= 26,800 \times 960.7$ B.t.u./lb.	$= 25,750,000$	100,000
Deduct for heating, 100,000 x (210° — 180° = 30°)	$= 3,000,000$		
Available for evaporation.....		22,750,000	
L @ 209° = 972.2 B.t.u./lb. E1 = $\frac{22,750,000}{972.2}$ =			23,400
Transferred to No. 2.....			76,600
(2) Vapor ex. No. 1.....		22,750,000	
Add flash, 76,600 x (210° — 185° = 25°).....		1,915,000	
Available for evaporation.....		24,665,000	
L @ 182° = 988.7 B.t.u./lb. E2 = $\frac{24,665,000}{988.7}$ =			24,900
Transferred to No. 3.....			51,700

TRIPLE EFFECT HEAT BALANCE—CYCLE "A" (Continued).

Bodies	Specifications	Heat B.t.u.'s	Juice, Pounds
(3) Vapor ex. No. 2.....		24,665,000	
Add flash, $51,700 \times (185^\circ - 134^\circ = 51^\circ)$		2,640,000	
Available for evaporation.....		27,305,000	
$L @ 125^\circ = 1,021.6 \text{ B.t.u./lb. } E_3 = \frac{27,305,000}{1,021.6} =$			26,700
Syrup ex. No. 3.....			25,000

To convert steam consumption to "from and at 212° ," multiply by the ratio of L at 227° divided by L at 212° thus:

$$\frac{L \text{ at } 227 = 960.7}{L \text{ at } 212 = 970.4} = 0.99.$$

Steam consumption from and at $212^\circ = 26,800 \text{ lbs.} \times 0.99 = 26,500 \text{ lbs.}$

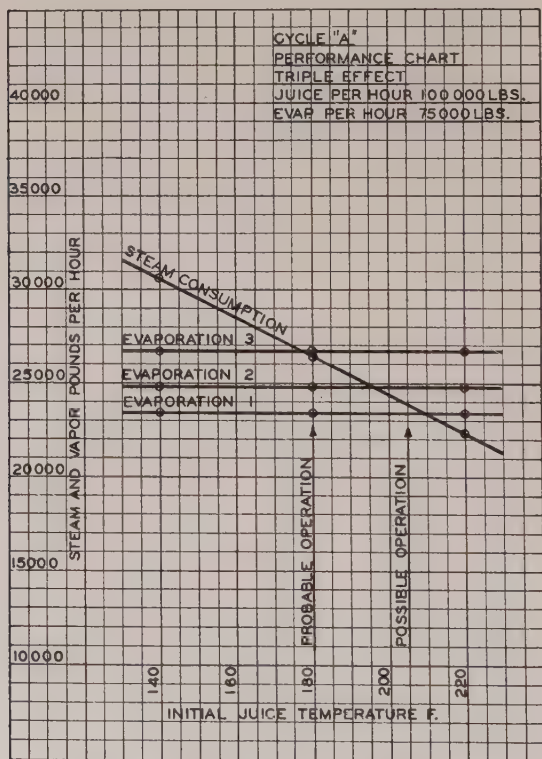


FIG. 4.

Cycle "B". In Cycle "B" the operation is in triple effect but the cold juices are heated from 100° to 200° by using vapors from the first body of the evaporator. The balance of the heating from 200° to 212°

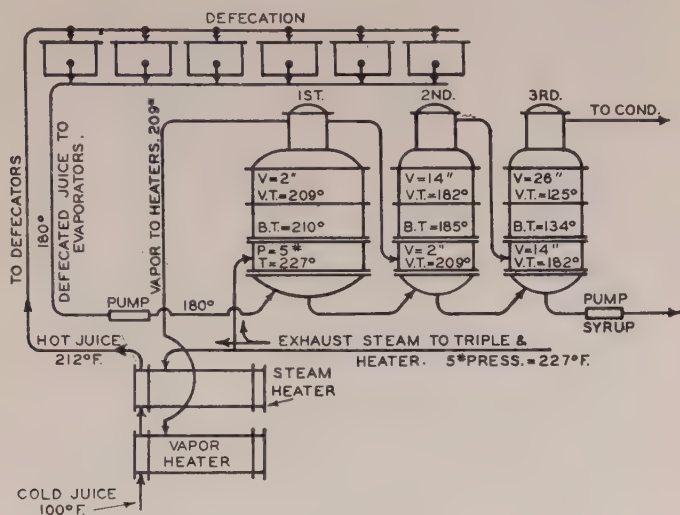


FIG. 5.—Cycle "B"—Triple Effect with Vapor and Exhaust Heater.

is done with either live or exhaust steam. This cycle is used only to a limited extent and is in reality an intermediate step to more modern equipment. We know of two such installations on a fair scale.

The distribution of steam is as follows:

	Pounds of Steam F. & A. 212° F.
(1) Power	3,940
(2) Exhaust juice heating $\frac{100,000 \times (212 - 200 = 12)}{970.4}$	1,235
(3) Evaporation and vapor heating	33,450
(4) Vacuum pans	20,000
(5) Radiation	2,500
Total	61,125
Available from bagasse alone (10 per cent. fibre)	54,000
Deficit to be made up with other fuel	7,125

If heat is conserved in the defecated juice and the feed to the evaporator is maintained at 205° F., instead of 180° F., as estimated, the consumption of the triple is reduced by 2570 lbs. per hour. The deficit is then 4555 lbs. from and at 212° F. (see curve).

TRIPLE EFFECT WITH VAPOR HEATER HEAT BALANCE—
CYCLE “B” (Continued).

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(3) Vapor ex. No. 2.....		21,240,000	
Add flash, 48,200 x (185° — 134° = 51°) =.....		2,460,000	
Available for evaporation.....		23,700,000	
L @ 125° = 1,021.6 B.t.u./lb. E ₃ = $\frac{23,700,000}{1,021.6}$ =.....			23,200
Syrup ex. No. 3.....			25,000
Steam from and at 212° = 33,800 x 0.99 = 33,450.			

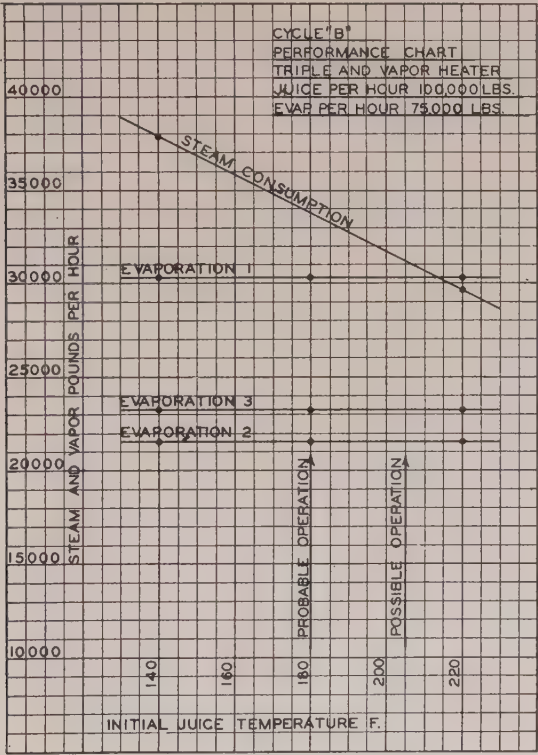


FIG. 7.

Cycle “C.” Cycle “C,” in which juice is heated with live or exhaust steam and evaporation is accomplished in a straight quadruple effect, fairly well represents the practice of a large number of sugar mills in Cuba and Porto Rico. It does not, however, result in a balance for the assumed conditions. The steam distribution is as follows :

		Pounds of Steam F. & A. 212° F.
(1)	Power	3,940
(2)	Heating	11,540
(3)	Evaporation	20,400
(4)	Vacuum pans	20,000
(5)	Radiation	2,500
(6)	Total	58,380
	Available from bagasse alone (10 per cent. fibre).....	54,000
	Deficit to be made up with other fuel.....	4,380

If the heat in the defecated juice is conserved and the feed reaches the quad. at 205° F. instead of 180° F., the steam consumption of the evaporator is reduced by 2570 lbs. per hour. The deficit will be 1810 lbs. per hour from and at 212° (see curve).

STANDARD QUADRUPLER HEAT BALANCE—CYCLE "C."

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(1)	Steam, 20600 lbs. @ 227° = 20,600 x 960.7 B.t.u./lb...	19,800,000	100,000
	Deduct for heating 100,000 x (214.5° — 180° = 34.5°) =	3,450,000	
	Available for evaporation.....	16,350,000	
	L @ 214° = 969.1 B.t.u./lb. $E_1 = \frac{16,350,000}{969.1} =$		16,900
	Transferred to No. 2.....		83,100
(2)	Vapor ex. No. 1.....	16,350,000	
	Add flash, 83,100 x (214.5° — 198° = 16.5°).....	1,370,000	
	Available for evaporation.....	17,720,000	
	L @ 197° = 979.7 B.t.u./lb. $E_2 = \frac{17,720,000}{979.7} =$		18,100
	Transferred to No. 3.....		65,000
(3)	Vapor ex. No. 2.....	17,720,000	
	Add flash, 65,000 x (198° — 175° = 23°) =.....	1,495,000	
	Available for evaporation.....	19,215,000	
	L @ 173° = 994 B.t.u./lb. $E_3 = \frac{19,215,000}{994} =$		19,350
	Transferred to No. 4.....		45,650
(4)	Vapor ex. No. 3.....	19,215,000	
	Add flash, 45,650 x (175° — 134° = 41°) =.....	1,870,000	
	Available for evaporation.....	21,085,000	
	L @ 125° = 1,021.6 B.t.u./lb. $E_4 = \frac{21,085,000}{1,021.6} =$		20,650
	Syrup ex. No. 4.....		25,000
	Steam from and at 212° = 20,600 x 0.99 = 20,400 lbs.		

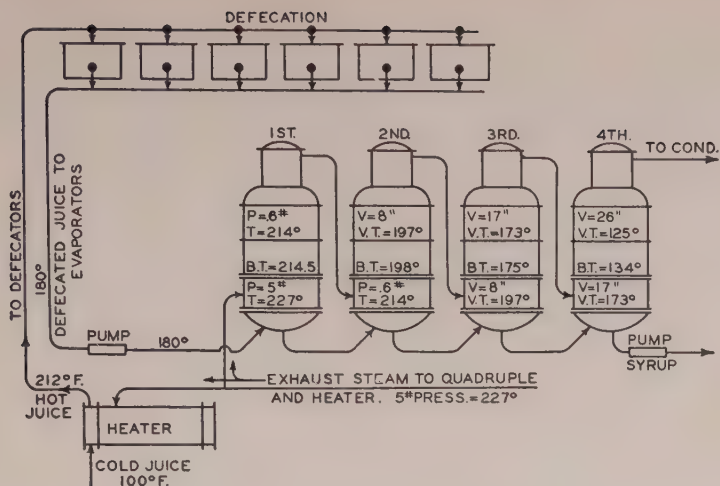


FIG. 8.—Cycle "C"—Quadruple Effect with Exhaust Heater.

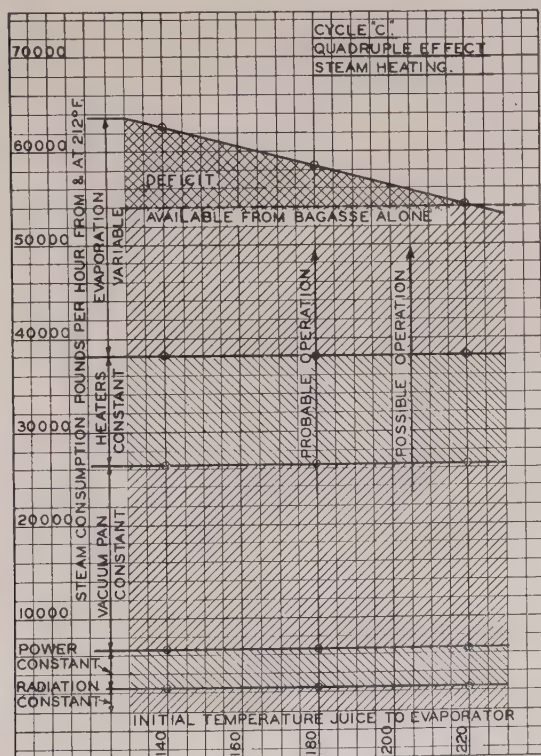


FIG. 9.

Cycle "D." Cycle "D" represents developments from Cycle "C." In this case the evaporation is accomplished in quadruple effect although the first body should have about twice the heating surface of the others.

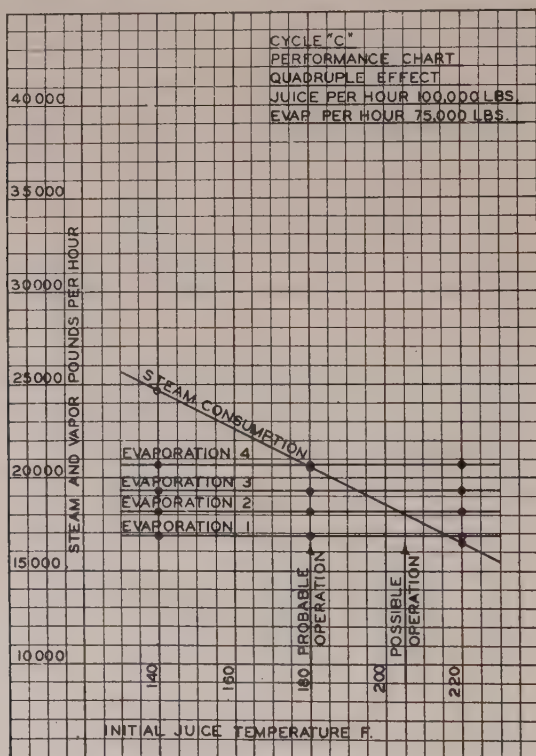


FIG. 10.

The cold juice is heated to 205° F. with vapors from the first effect at 212° F. The remainder of the heating operation is done with live or exhaust steam. The heat distribution is as follows:

	Pounds of Steam F. & A. 212° F.
(1) Power	3,940
(2) Heating (from 205° to 212°).....	720
(3) Evaporation in quad. and vapor heating.....	28,700
(4) Vacuum pans	20,000
(5) Radiation losses	2,500
(6) Total	55,860
Available from bagasse alone (10 per cent. fibre).....	54,000
Deficit	1,860

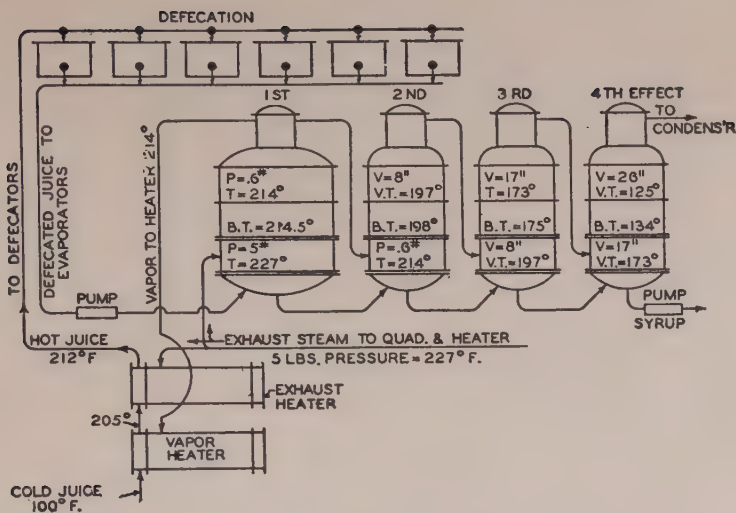


FIG. 11.—Cycle "D"—Quadruple Effect with Vapor and Exhaust Heaters.

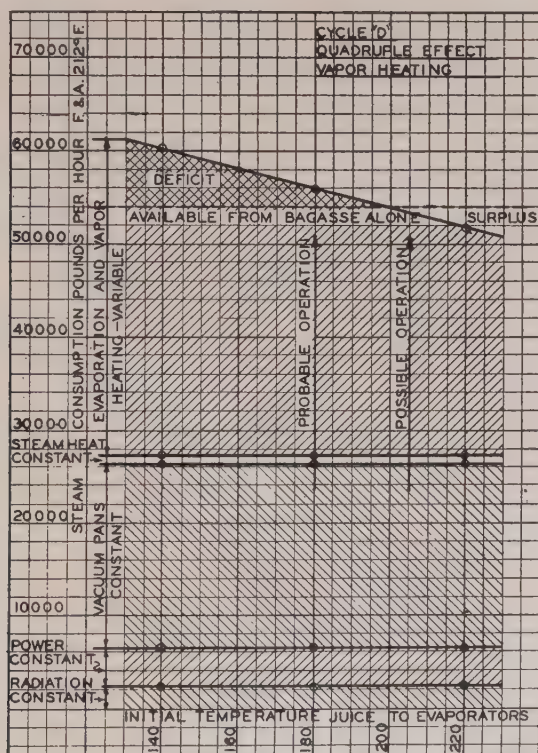


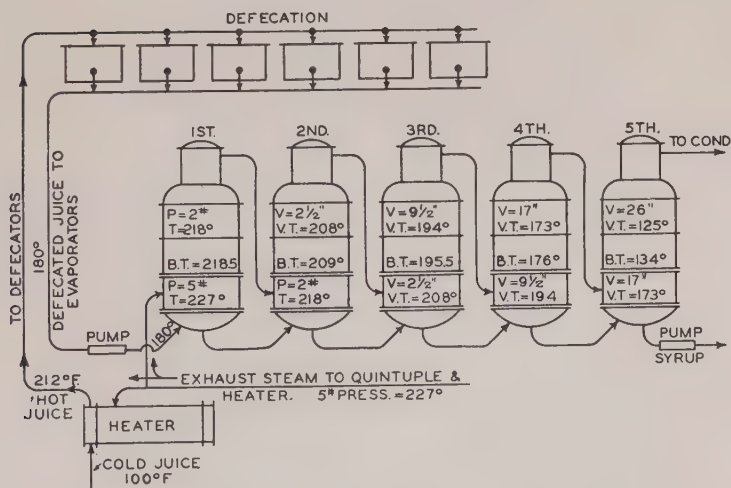
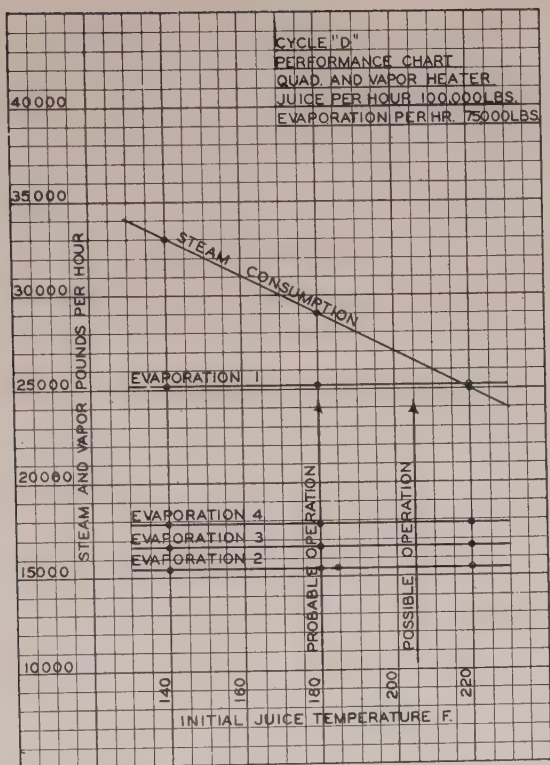
FIG. 12.

With the conservation of hot juice at 205° F., the steam consumption of the quadruple is reduced 2570 lbs. In this case there is an excess of 710 lbs. per hour from and at 212° F. (see diagram).

QUADRUPLE WITH VAPOR HEATING—CYCLE "D."

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(1)	29,000 lbs. @ 227° = 29,000 x 960.7 B.t.u./lbs. =	27,850,000	100,000
	Deduct for heating 100,000 x (214.5° — 180° = 34.5°) =	3,450,000	
	Available for evaporation.	24,400,000	
	L @ 214° = 969.1 B.t.u./lb. $E_1 = \frac{24,400,000}{969.1} =$		25,200
	Transferred to No. 2.		74,800
(2)	Vapor ex. No. 1.	24,400,000	
	Add flash, 74,800 x (214.5° — 198° = 16.5°) =	1,234,000	
	Available total heat.	25,634,000	
	Deduct for heater, 100,000 x (205° — 100° = 105°) =	10,500,000	
	Available for evaporation.	15,134,000	
	L @ 197° = 979.7 B.t.u./lb. $E_2 = \frac{15,134,000}{979.7} =$		15,450
	Transferred to No. 3.		59,350
(3)	Vapor ex. No. 2.	15,134,000	
	Add flash, 59,350 x (198° — 175° = 23°) =	1,365,000	
	Available for evaporation.	16,499,000	
	L @ 173° = 994 B.t.u./lb. $E_3 = \frac{16,499,000}{994} =$		16,600
	Transferred to No. 4.		42,750
(4)	Vapor ex. No. 3.	16,499,000	
	Add flash, 42,750 x (175° — 134° = 41°) =	1,750,000	
	Available for evaporation.	18,249,000	
	L @ 125° = 1021.6 B.t.u./lb. $E_4 = \frac{18,249,000}{1021.6} =$		17,850
	Syrup ex. No. 4.		24,900
	Steam from and at 212° = 29,000 x 0.99 = 28,700 lbs.		

Cycle "E." The use of the straight quintuple effect, represented by Cycle "E," is a recent development in the Cuban practice dating back perhaps five years. Its performance has been very satisfactory, more so than Cycle "D," as the operation is not subject to violent fluctuations, such as occur when there is a temporary mill stoppage. This will be discussed in more detail later.



The steam distribution is as follows:

	Pounds of Steam F. & A. 212° F.
(1) Power	3,940
(2) Heating	11,540
(3) Evaporation	17,150
(4) Vacuum pans	20,000
(5) Radiation	2,500
(6) Total	55,130
Available from combustion of bagasse alone (10 per cent.)	54,000
Deficit	1,130

It will be noted that this cycle practically represents a balance. The reduction in steam consumption, with hot juice conservation to 205° as before, is 2570 lbs. This results in an available excess steam supply of 1440 lbs. per hour from and at 212° F. (see diagram).

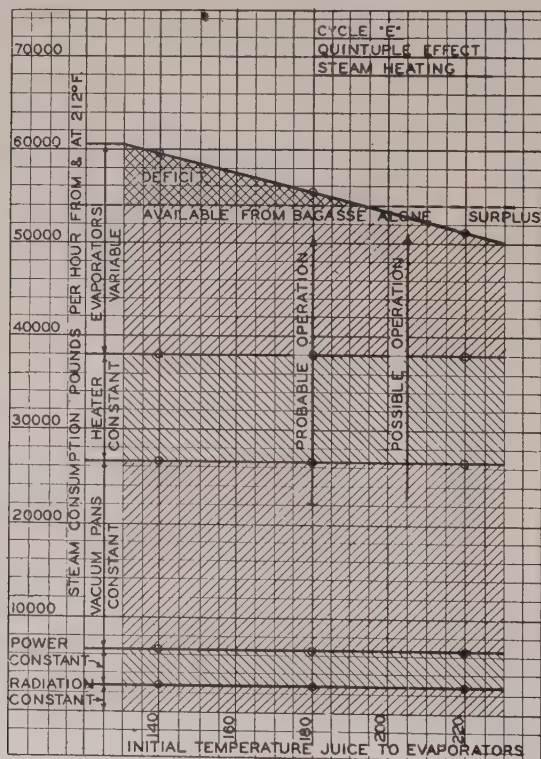
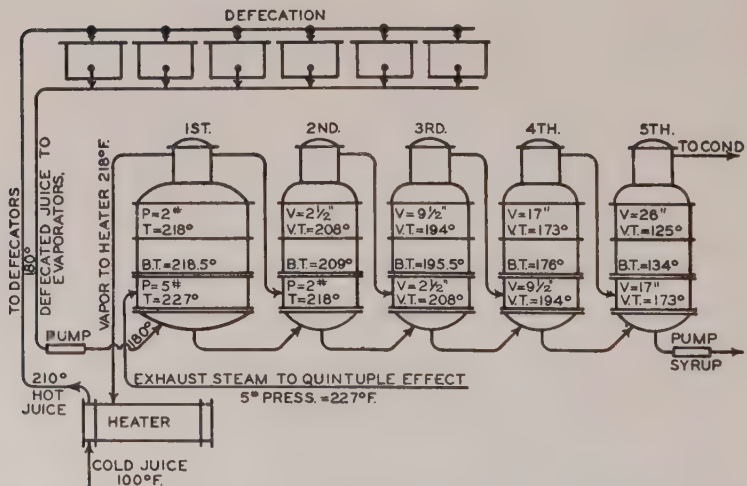
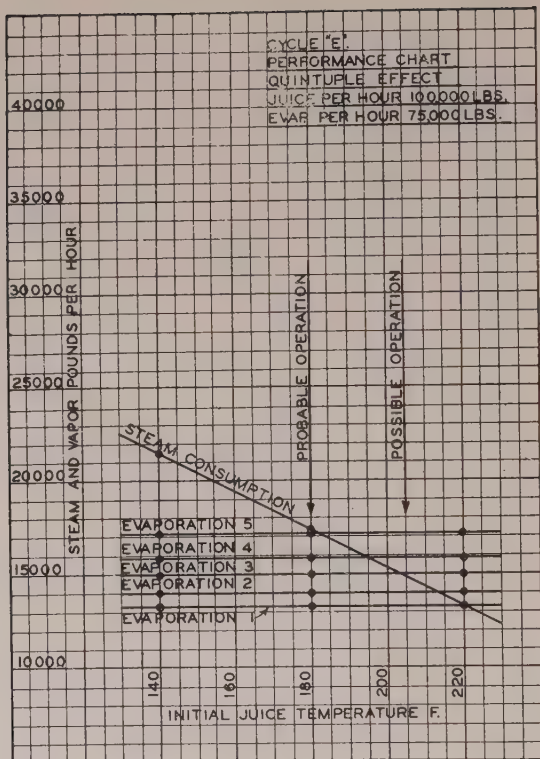


FIG. 15.

STANDARD QUINTUPLE HEAT BALANCE—CYCLE "E."

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(1)	17,350 lbs. @ 227° = $17,350 \times 960.7$ B.t.u./lb. =	16,650,000	100,000
	Deduct for heating, $100,000 \times (218.5^\circ - 180^\circ = 38.5^\circ)$ =	3,850,000	
	Available for evaporation.....	12,800,000	
	L @ 218° = 966.5 B.t.u./lb. $E_1 = \frac{12,800,000}{966.5}$ =		13,250
	Transferred to No. 2.....		86,750
(2)	Vapor ex. No. 1.....	12,800,000	
	Add flash, $86,750 \times (218.5^\circ - 209^\circ = 9.5^\circ)$ =	825,000	
	Available for evaporation.....	13,625,000	
	L @ 208° = 972.9 B.t.u./lb. $E_2 = \frac{13,625,000}{972.9}$ =		14,000
	Transferred to No. 3.....		72,750
(3)	Vapor ex. No. 2.....	13,625,000	
	Add flash, $72,750 \times (209^\circ - 195.5^\circ = 13.5^\circ)$ =	982,000	
	Available for evaporation.....	14,607,000	
	L @ 194° = 981.5 B.t.u./lb. $E_3 = \frac{14,607,000}{981.5}$ =		14,870
	Transferred to No. 4.....		57,880
(4)	Vapor ex. No. 3.....	14,607,000	
	Add flash, $57,880 \times (195.5^\circ - 176^\circ = 19.5^\circ)$ =	1,130,000	
	Available for evaporation.....	15,737,000	
	L @ 173° = 994 B.t.u./lb. $E_4 = \frac{15,737,000}{994}$ =		15,830
	Transferred to No. 5.....		42,050
(5)	Vapor ex. No. 4.....	15,737,000	
	Add flash, $42,050 \times (176^\circ - 134^\circ = 42^\circ)$ =	1,765,000	
	Available for evaporation.....	17,502,000	
	L @ 125° = $1,021.6$ B.t.u./lb. $E_5 = \frac{17,502,000}{1,021.6}$ =		17,100
	Syrup ex. No. 5.....		24,950
	Steam from and at 212° = $17,350 \times 0.99$ = 17,150 lbs.		

Cycle "F." Cycle "F," which, we believe, has not been used extensively in the cane sugar industry, is discussed here only as a matter of interest and to show that, if further economy is necessary, it is easily possible to secure substantial results by elaborating slightly on our established practice. A quintuple effect with a large first body would be used so as to supply enough vapor at 218° F. to heat the juices to 210° F., a point sufficiently high for defecation.



The heat distribution would be as follows :

	Pounds of Steam F. & A. 212° F.
(1) Power	3,940
(2) Heating	0
(3) Quintuple with complete vapor heating to 210°	26,350
(4) Vacuum pans	20,000
(5) Radiation losses	2,500
(6) Total	52,790
Available from bagasse alone (10 per cent. fibre)	54,000
Excess	1,210

The steam consumption will be reduced by 2,570 lbs. if heat is conserved in defecated juice to 205° F. This increases our available excess to 3,780 lbs.

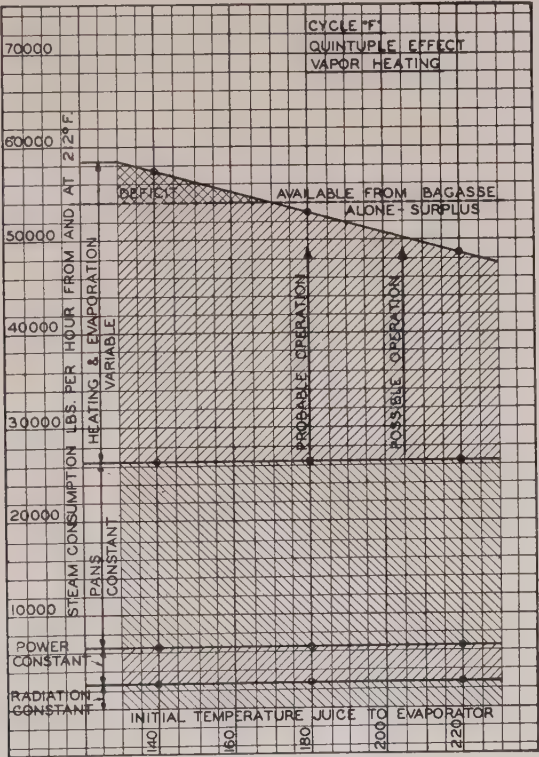


FIG. 18.

QUINTUPLE EFFECT WITH VAPOR HEAT—CYCLE "F."

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(1)	Steam, 26,600 lbs. @ $227^{\circ} = 26,600 \times 960.7$ B.t.u./lb. =	25,550,000	100,000
	Deduct for heating, $100,000 \times (218.5^{\circ} - 180^{\circ} = 38.5^{\circ}) =$	3,850,000	
	Available for evaporation.....	21,700,000	
	L @ $218^{\circ} = 966.5$ B.t.u./lb. $E_1 = \frac{21,700,000}{966.5} =$		22,500
	Transferred to No. 2.....		77,500
(2)	Vapor ex. No. 1.....	21,700,000	
	Add flash, $77,500 \times (218.5^{\circ} - 209^{\circ} = 9.5^{\circ}) =$	737,000	
	Available heat	22,437,000	
	Deduct for vap. heat, $100,000 \times (210^{\circ} - 100^{\circ} = 110^{\circ}) =$	11,000,000	
	Available for evaporation.....	11,437,000	
	L @ $208^{\circ} = 972.9$ B.t.u./lb. $E_2 = \frac{11,437,000}{972.9} =$		11,770
	Transferred to No. 3.....		65,730
(3)	Vapor ex. No. 2.....	11,437,000	
	Add flash, $65,730 \times (209^{\circ} - 195.5^{\circ} = 13.5^{\circ}) =$	887,000	
	Available for evaporation.....	12,324,000	
	L @ $194^{\circ} = 981.5$ B.t.u./lb. $E_3 = \frac{12,324,000}{981.5} =$		12,570
	Transferred to No. 4.....		53,160
(4)	Vapor ex. No. 3.....	12,324,000	
	Add flash, $53,160 \times (195.5^{\circ} - 176^{\circ} = 19.5^{\circ}) =$	1,036,000	
	Available for evaporation.....	13,360,000	
	L @ $173^{\circ} = 994$ B.t.u./lb. $E_4 = \frac{13,360,000}{994} =$		13,450
	Transferred to No. 5.....		39,710
(5)	Vapor ex. No. 4.....	13,360,000	
	Add flash, $39,710 \times (176^{\circ} - 134^{\circ} = 42^{\circ}) =$	1,667,000	
	Available for evaporation.....	15,027,000	
	L @ $125^{\circ} = 1021.6$ B.t.u./lb. $E_5 = \frac{15,027,000}{1,021.6} =$		14,700
	Syrup ex. 5		25,010
	Steam from and at $212^{\circ} = 26,600 \times 0.99 = 26,350$ lbs.		

Cycle "G." The cycles previously discussed are fairly simple and very easily analyzed and understood. It is quite different with cycle "G," which consists of a plain quadruple operating as in cycle "C," ahead of which has been placed a single body or "Pauly." The "Pauly," which is named after the man who first used this method, operates entirely with live steam. Its vapors are discharged either into the exhaust main or

into the steam belt of the first effect, in which case they mix with the exhaust, thus giving virtually the same result. The juice coming from the defecation is fed direct into this "Pauly" and thence to the quadruple effect, where it is eventually concentrated to the proper density.

This operation is difficult to analyze, as the work of the "Pauly" is subject to wide fluctuations, its main function being to act as a pressure-reducing valve between the live and exhaust steam line; and as the exhaust

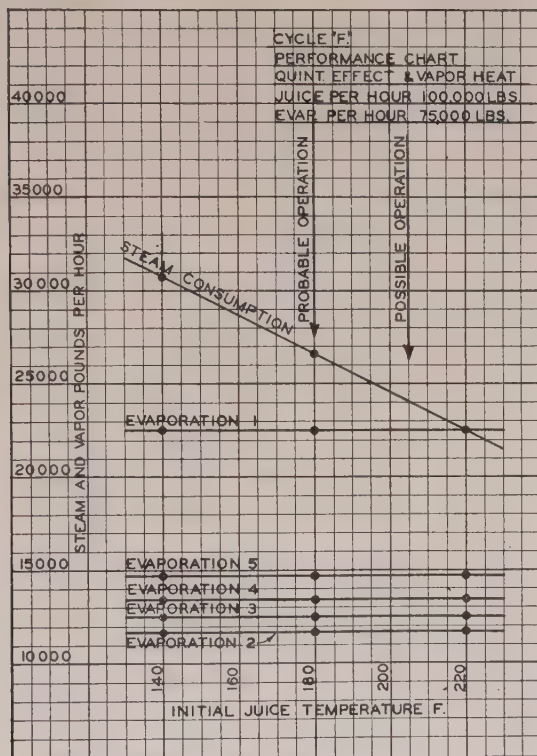


FIG. 19.

varies in supply and demand its work must vary accordingly. These variations directly affect the economy of the entire operation. It is, therefore, necessary to make certain assumptions upon which to predicate our results and the calculations involved are rather complicated. Ordinarily in making such calculations many engineers assume that for every pound of steam condensed in the "Pauly" there is one pound of vapor produced. This assumption is far from being correct.

In the particular case under consideration before any evaporation occurs, it is necessary to heat the entire quantity of juice (100,000 lbs.) from its temperature as it enters the equipment (here assumed at 180° F.)

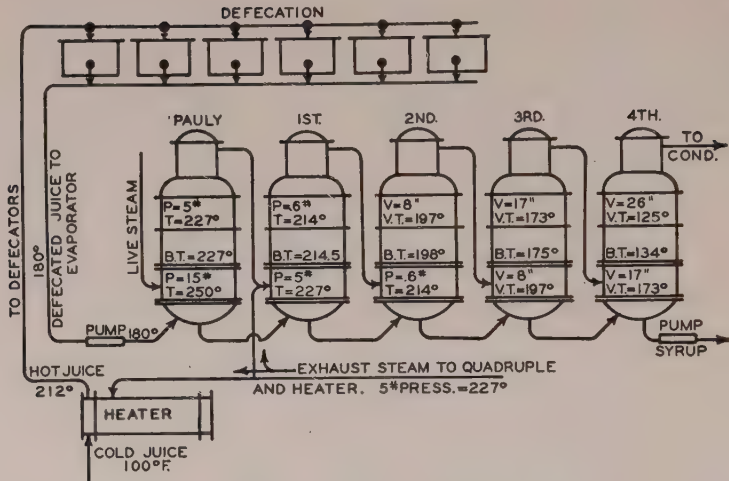


FIG. 20.—Cycle "G"—Quadruple Effect with Pauly and Exhaust Heater.

to the temperature at which ebullition will take place (227° F. in this case), a total rise of 47° F. This requires 4,700,000 B.t.u.'s. It is customary to carry a pressure of 15 to 20 lbs. in the calandria of the "Pauly." The latent heat at 15 lbs. being 945.3, the amount of steam required to make up this 4,700,000 B.t.u.'s will be 4965 lbs., corresponding to 4830 lbs. from and at 212° F. *There is nothing gained in economy over a plain quadruple effect until after this point is passed; and, as far as economy is concerned, this heat could just as well have been put directly into the quadruple.*

It is readily seen, therefore, that the advantage in operating the "Pauly" is obtained only in case of a large proportion of live to exhaust steam. This is a factor depending directly upon the steam economy of power units as well as the power consumption of the factory in proportion to its capacity.

We previously assumed that it would require 1500 actual horsepower to operate this particular plant. If we assume 30 lbs. per H.P. hour, the amount of exhaust so produced would be 45,000 lbs. per hour. The available supply from bagasse alone is given as 54,000 lbs.

Roughly, we have as follows:

	Pounds of Steam F. & A. 212° F.	
Available steam from bagasse alone.....	54,000	
Made into exhaust by power generators	45,000	
Maximum available for use in "Pauly".....	9,000	9,000
To overcome useless range	4,830	
Available for useful range of "Pauly".....	4,170	
Probable shortage requiring extra fuel.....		3,000
Total available for "Pauly".....		12,000

The steam consumption of the Plant would be :

	Pounds of Steam F. & A. 212° F.
(1) Power	3,940
(2) Heating	11,540
(3) Evaporation	18,700
(4) Vacuum pans	20,000
(5) Radiation losses	2,500
(6) Total	56,680
Available steam supply bagasse alone.....	54,000
Deficit	2,680

If the heat in defecated juice is conserved to 205° F. as before, the steam consumption of the evaporating station will be reduced about 3000 lbs., leaving an excess of 320 lbs.

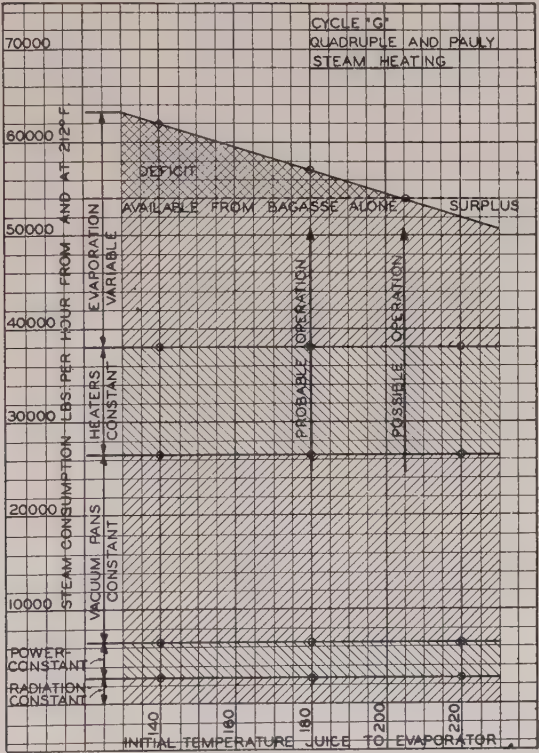


FIG. 21.

It is, therefore, evident that this cycle is the inferior of "D," "E" and "F" under all conditions.

At our assumed operating point it is worthy of consideration to note

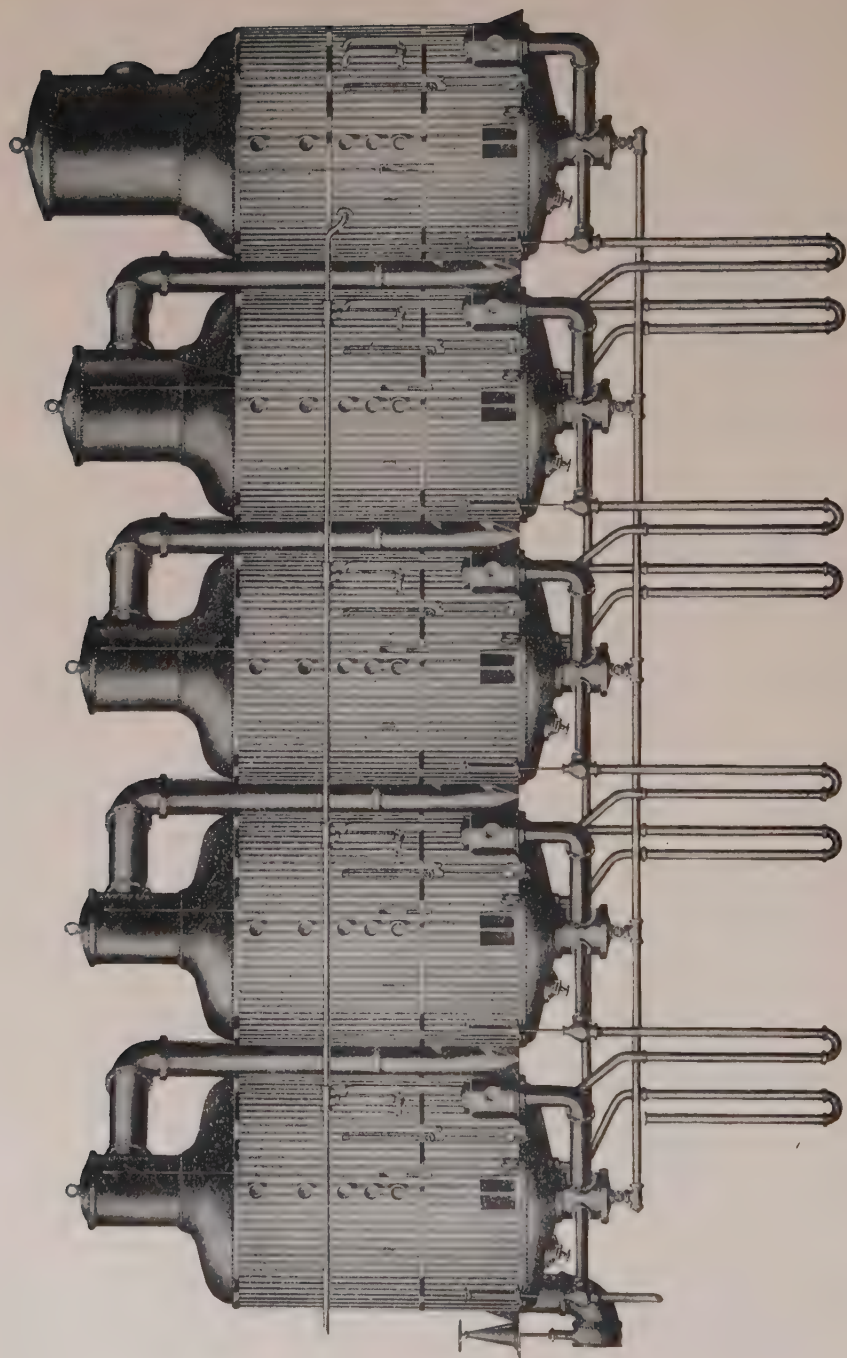


FIG. 21a.

(Courtesy U.S.C.I. Pipe & Fdy. Co.)

that the steam consumption of the "Pauly" is approximately the same as that of the quad following it.

It is also interesting to note that if the steam consumption of the "Pauly" is increased to 17,500 lbs. per hour from and at 212° F. that its evaporation is equal to the steam consumption of the quad. This results in a quintuple operating entirely on live steam.

The data on which these conclusions are based is discussed in detail in the following pages.

Intermediate Calculations for Cycle "G."

In order to establish the performances of the "Pauly" and quadruple operating jointly, we have calculated heat balances for the quadruple alone under various "Pauly" performances. Heat Balance G-1 shows the operation of the quad if the "Pauly" is used merely to heat the juices up to the temperature corresponding to the pressure on its liquor side. This pressure is assumed at 5 lbs. and the temperature is, therefore, 227° F. The quad steam consumption is 15,550 lbs. per hour from and at 212° F. for the total work contemplated; namely, an evaporation of 75,000 lbs. from 100,000 lbs. of juice.

G-2 shows the heat balance of the quad alone when the "Pauly" is evaporating 5000 lbs. per hour in addition to the heating operation. The steam consumption of the quad under these conditions is 14,450 lbs. from and at 212° F.

G-3 is exactly similar but the "Pauly" evaporates 10,000 lbs. instead of 5000 lbs. as above and the steam consumption of the quad becomes 13,350 lbs. from and at 212° F.

G-4 shows the conditions when the "Pauly" evaporates 15,000 lbs. per hour. Here, the quad uses 12,220 lbs. of steam from and at 212° F.

STANDARD QUAD. HEAT BALANCE—CYCLE G-1.

WITH "PAULY" HEATING JUICE TO 227° : (NO EVAP.).

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(1) Steam 15,700 lbs. @ 227° = $15,700 \times 960.7$ B.t.u./lb. =		15,100,000	100,000
Add juice flash, 100,000 x ($227^{\circ} - 214.5^{\circ} = 12.5^{\circ}$) =		1,250,000	
Available for evaporation.....		16,350,000	
L @ 214° = 969.1 B.t.u./lb. $E_1 = \frac{16,350,000}{969.1} =$			16,900
Transferred to No. 2			83,100
(2) Vapor ex. (1).....		16,350,000	
Add liquid flash, 83,100 x ($214.5^{\circ} - 198^{\circ} = 16.5^{\circ}$) =		1,370,000	
Available for evaporation.....		17,720,000	
L @ 197° = 979.7 B.t.u./lb. $E_2 = \frac{17,720,000}{979.7} =$			18,100
Transferred to No. 3.....			65,000

STANDARD QUAD. HEAT BALANCE—CYCLE G-1.
WITH "PAULY" HEATING JUICE TO 227°: (No EVAP.). (Continued).

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(3) Vapor ex. (2).....		17,720,000	
Add liquid flash, 65,000 x (198° — 175° = 23°) =.....		1,495,000	
Available for evaporation.....		19,215,000	
L @ 173° = 994 B.t.u./lb. E ₃ = $\frac{19,215,000}{994}$ =.....			19,350
Transferred to No. 4.....			45,650
(4) Vapor ex. (3).....		19,215,000	
Add liquid flash, 45,650 x (175° — 134° = 41°) =.....		1,870,000	
Available for evaporation.....		21,085,000	
L @ 125° = 1,021.6 B.t.u./lb. E ₄ = $\frac{21,085,000}{1,021.6}$ =.....			20,650
Syrup ex. No. 4.....			25,000
Steam from and at 212° F. = 15,700 x 0.99 = 15,550 lbs.			

STANDARD QUAD. HEAT BALANCE—CYCLE G-2.
WITH "PAULY" EVAPORATING 5,000 LBS. PER HR.

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(1) Steam 14,600 lbs. @ 227° = 14,600 x 960.7 B.t.u./lb. =		14,030,000	95,000
Add flash, 95,000 x (227° — 214.5° = 12.5°) =.....		1,188,000	
Available for evaporation.....		15,218,000	
L @ 214° = 969.1 B.t.u./lb. E ₁ = $\frac{15,218,000}{969.1}$ =.....			15,700
Transferred to No. 2.....			79,300
(2) Vapor ex. No. 1.....		15,218,000	
Add flash, 79,300 x (214.5° — 198° = 16.5°) =.....		1,310,000	
Available for evaporation.....		16,528,000	
L @ 197° = 979.7 B.t.u./lb. E ₂ = $\frac{16,528,000}{979.7}$ =.....			16,850
Transferred to No. 3.....			62,450
(3) Vapor ex. No. 2.....		16,528,000	
Add flash, 62,450 x (198° — 175° = 23°) =.....		1,436,000	
Available for evaporation.....		17,964,000	
L @ 173° = 994 B.t.u./lb. E ₃ = $\frac{17,964,000}{994}$ =.....			18,050
Transferred to No. 4.....			44,400
(4) Vapor ex. No. 3.....		17,964,000	
Add flash, 44,400 x (175° — 134° = 41°) =.....		1,820,000	
Available for evaporation.....		19,784,000	
L @ 125° = 1,021.6 B.t.u./lb. E ₄ = $\frac{19,784,000}{1,021.6}$ =.....			19,400
Syrup ex. No. 4.....			25,000
Steam from and at 212° F. = 14,600 x 0.99 = 14,450 lbs.			

THE HEAT BALANCE IN THE CANE SUGAR FACTORY 273

STANDARD QUAD. HEAT BALANCE—CYCLE G-3.

"PAULY" EVAPORATING 10,000 LBS.

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(1)	13,500 lbs. @ $227^{\circ} = 13,500 \times 960.7$ B.t.u./lb. =	12,960,000	90,000
	Add flash, $90,000 \times (227^{\circ} - 214.5^{\circ} = 12.5^{\circ}) =$	1,125,000	
	Available for evaporation.....	14,085,000	
	L @ $214^{\circ} = 969.1$ B.t.u./lb. $E_1 = \frac{14,085,000}{969.1} =$		14,520
	Transferred to No. 2.....		75,480
(2)	Vapor ex. No. 1.....	14,085,000	
	Add flash, $75,480 \times (214.5^{\circ} - 198^{\circ} = 16.5^{\circ}) =$	1,244,000	
	Available for evaporation.....	15,329,000	
	L @ $197^{\circ} = 979.7$ B.t.u./lb. $E_2 = \frac{15,329,000}{979.7} =$		15,640
	Transferred to No. 3.....		59,840
(3)	Vapor ex. No. 2.....	15,329,000	
	Add flash, $59,840 \times (198^{\circ} - 175^{\circ} = 23^{\circ}) =$	1,377,000	
	Available for evaporation.....	16,706,000	
	L @ $173^{\circ} = 994$ B.t.u./lb. $E_3 = \frac{16,706,000}{994} =$		16,800
	Transferred to No. 4.....		43,040
(4)	Vapor ex. No. 3.....	16,706,000	
	Add flash, $43,040 \times (175^{\circ} - 134^{\circ} = 41^{\circ}) =$	1,764,000	
	Available for evaporation.....	18,470,000	
	L @ $125^{\circ} = 1021.6$ B.t.u./lb. $E_4 = \frac{18,470,000}{1,021.6} =$		18,040
	Syrup ex. No. 4.....		25,000
	Steam from and at 212° F. = $13,500 \times 0.99 = 13,350$ lbs.		

STANDARD QUAD. HEAT BALANCE—CYCLE G-4.

WITH "PAULY" EVAPORATION 15,000 LBS. PER HOUR.

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(1)	Steam, 12,360 lbs. @ $227^{\circ} = 12,360 \times 960.7$ B.t.u./lb. =	11,880,000	85,000
	Add flash, $85,000 \times (227^{\circ} - 214.5^{\circ} = 12.5^{\circ}) =$	1,062,000	
	Available for evaporation.....	12,942,000	
	L @ $214^{\circ} = 969.1$ B.t.u./lb. $E_1 = \frac{12,942,000}{969.1} =$		13,330
	Transferred to No. 2.....		71,670
(2)	Vapor ex. No. 1.....	12,942,000	
	Add flash, $71,670 \times (214.5^{\circ} - 198^{\circ} = 16.5^{\circ}) =$	1,182,000	
	Available for evaporation.....	14,124,000	
	L @ $197^{\circ} = 979.7$ B.t.u./lb. $E_2 = \frac{14,124,000}{979.7} =$		14,400
	Transferred to No. 3.....		57,270

STANDARD QUAD. HEAT BALANCE—CYCLE G-4.
WITH "PAULY" EVAPORATION 15,000 LBS. PER HOUR. (Continued).

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(3) Vapor ex. No. 2.....		14,124,000	
Add flash, 57,270 x (198° — 175° = 23°) =		1,317,000	
Available for evaporation.....		15,441,000	
L @ 173° = 994 B.t.u./lb. E ₃ = $\frac{15,441,000}{994}$ =			15,530
Transferred to No. 4			41,740
(4) Vapor ex. No. 3.....		15,441,000	
Add flash, 41,740 x (175° — 134° = 41°) =		1,710,000	
Available for evaporation		17,151,000	
L @ 125° = 1021.6 B.t.u./lb. E ₄ = $\frac{17,151,000}{1,021.6}$ =			16,760
Syrup ex. No. 4.....			24,980
Steam from and at 212° F. = 12,360 x 0.99 = 12,220 lbs.			

Total Set Performance.

The calculations and chart G-x show the total steam consumption for the set, assuming the initial juice feed temperature at 205° F. Four points have been located on the curve: The first, when the "Pauly" merely brings the juice to boiling as in heat balance G-1; the second, with the "Pauly" evaporating 5000 lbs. per hour as in G-2; the third, with the "Pauly" evaporating 10,000 lbs. as in G-3, and the fourth with the "Pauly" evaporating 15,000 lbs. as in G-4. The information developed is plotted on the chart G-x ordinates representing steam consumption in pounds per hour from and at 212° F. and also the economy of the set over a plain quadruple operating under the same conditions. The evaporation in the "Pauly" is also shown as ordinates. Abscissae represent the steam consumption of the "Pauly" alone in pounds per hour from and at 212° F.

In the same way G-y and the corresponding chart represent the same information except that the initial temperature of the juice is taken at 180° F. instead of 205° as above. G-Z represents similar information with the initial juice temperature at 140° F.

SUMMARY OF "PAULY" CALCULATIONS.

CHART G-X.

205° Feed

(1) Heating Effect

100,000 lbs. from (205° to 227° = 22°)

22° x 100,000 lbs. = 2,200,000 B.t.u.

Steam F & A 212° = $\frac{2,200,000}{970.4}$ = 2,265 lbs.

Steam Expense With "Pauly" Evap. 0—

Pauly 2,265 lbs.

Quad. Ex. G-1 15,550 "

Total 17,815 "

SUMMARY OF "PAULY" CALCULATIONS.

CHART G-X (Continued).

Steam Expense With "Pauly" Evap. 5000 lbs.		
Pauly	2,265	lbs.
Quad. Ex. G-2	14,450	"
Total	16,715	"
Steam Expense With "Pauly" Evap. 10,000 lbs.		
Pauly	2,265	lbs.
Quad. Ex. G-3	13,350	"
Total	15,615	"
Steam Expense With "Pauly" Evap. 15,000 lbs.		
Pauly	2,265	lbs.
Quad. Ex. G-4	12,220	"
Total	14,485	"

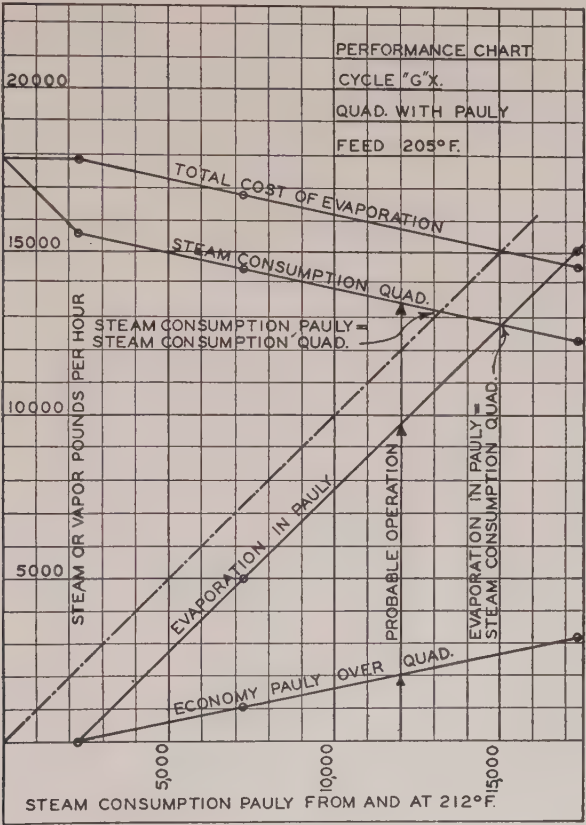


FIG. 22.

SUMMARY OF "PAULY" CALCULATIONS.

CHART G-Y.

180° Feed

(1) Heating Effect

100,000 lbs. from 180° to 227° = 47°.

47° x 100,000 lbs. = 4,700,000 B.t.u.

Steam F & A 212° = $\frac{4,700,000}{970.4}$ =	4,830 lbs.
--	------------

Steam Expense With "Pauly" Evap. 0—

Pauly	4,830 lbs.
-------------	------------

Quad. Ex. G-1	15,550 "
---------------------	----------

Total	20,380 "
-------------	----------

Steam Expense With "Pauly" Evap. 5000 lbs.

Pauly	4,830 lbs.
-------------	------------

Quad. Ex. G-2	14,450 "
---------------------	----------

Total	19,280 "
-------------	----------

Steam Expense With "Pauly" Evap. 10,000 lbs.

Pauly	4,830 lbs.
-------------	------------

Quad. Ex. G-3	13,350 "
---------------------	----------

Total	18,180 "
-------------	----------

Steam Expense With "Pauly" Evap. 15,000 lbs.

Pauly	4,830 lbs.
-------------	------------

Quad. Ex. G-4	12,220 "
---------------------	----------

Total	17,050 "
-------------	----------

SUMMARY OF "PAULY" CALCULATIONS.

CHART G-Z.

140° Feed

(1) Heating Effect

100,000 lbs. from 140° to 227° = 87°.

87 x 100,000 = 8,700,000 B.t.u.

Steam from and at 212° = $\frac{8,700,000}{970.4}$ =	8,960 lbs.
--	------------

Steam Expense With "Pauly" Evap. 0—

Pauly	8,960 lbs.
-------------	------------

Quad. Ex. G-1	15,500 "
---------------------	----------

Total	24,460 "
-------------	----------

Steam Expense With "Pauly" Evap. 5000 lbs.

Pauly	8,960 lbs.
-------------	------------

Quad. Ex. G-2	14,450 "
---------------------	----------

Total	23,410 "
-------------	----------

Steam Expense With "Pauly" Evap. 10,000 lbs.

Pauly	8,960 lbs.
-------------	------------

Quad. Ex. G-3	13,350 "
---------------------	----------

Total	22,310 "
-------------	----------

Steam Expense With "Pauly" Evap. 15,000 lbs.

Pauly	8,960 lbs.
-------------	------------

Quad. Ex. G-4	12,220 "
---------------------	----------

Total	21,180 "
-------------	----------

Composite Chart

We also submit a composite chart made up of the charts G-x, G-y and G-z in which is shown only the total steam expense of the entire set as ordinates and the steam consumption of the "Pauly" above as abscissae. We have shown in dotted lines interpolations for initial juice temperatures other than those assumed in G-x, G-y and G-z. The area of the chart is

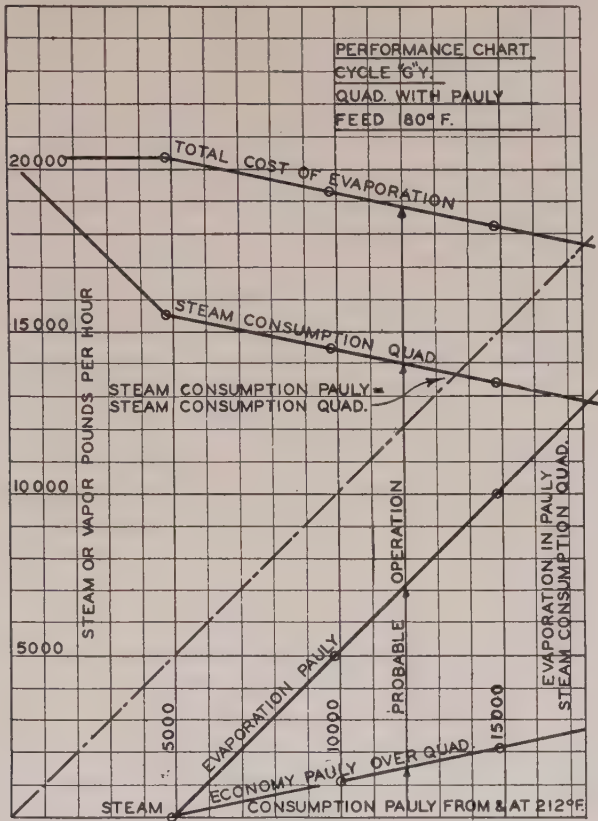
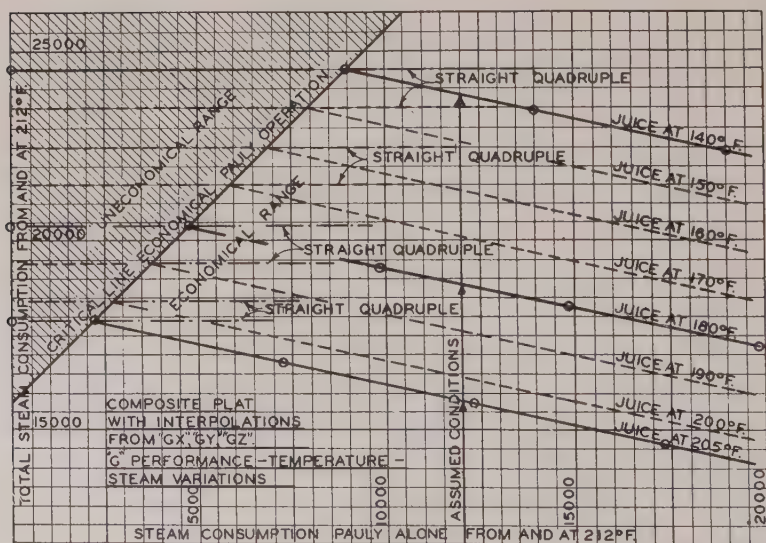
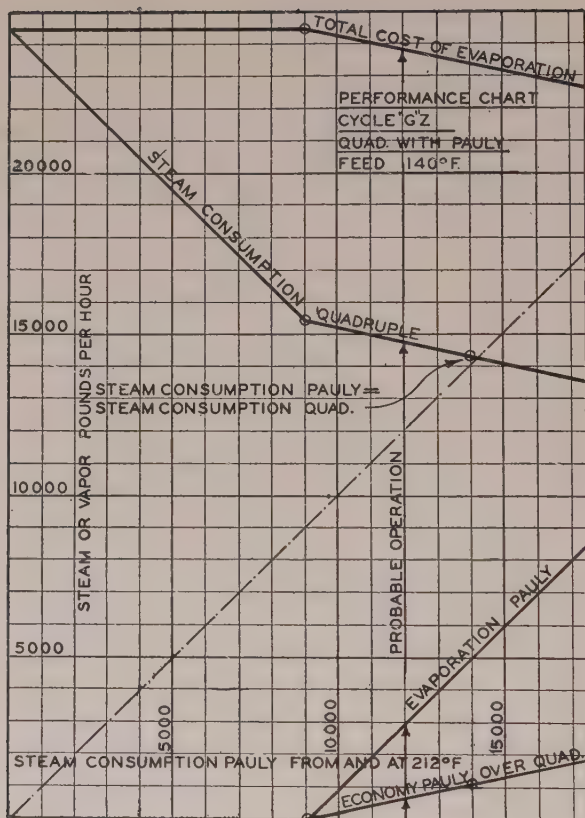


FIG. 23.

divided by the critical line, on the right of which the "Pauly" becomes economical; and on the left of which it ceases to be so.

On the right of the critical line, the depression from the horizontal shows the actual amount of steam saved over and above plain quadruple effect operation for any selected condition within the range of the chart.

We have also indicated the probable operating conditions taken here as a total steam consumption of 12,000 lbs. per hour from and at 212° F. for the "Pauly" alone.



Intermediate Distribution in Quadruple

Chart "G-I" is plotted with evaporation in the "Pauly" as abscissae and with ordinates representing the steam consumption of the quad alone and the evaporation in each body of the quad. The data for this chart is taken from heat balances G-1, G-2, G-3 and G-4. By referring to charts G-x, G-y and G-z, which represent the "Pauly" performances with initial juice feed temperatures of 205°, 180° and 140° F., the amount of evaporation in the "Pauly" alone, when using a total of 12,000 lbs. of steam from and at 212° F. per hour, can be determined. Chart G-I can now be used to

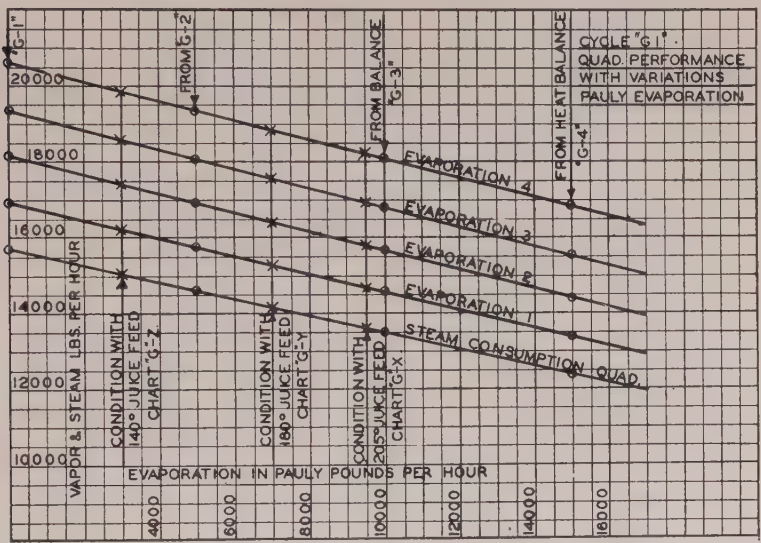


FIG. 26.

determine the steam consumption of the quad as well as the evaporation in each body for these conditions.

This information is plotted on the curves identified as cycle "G" which correspond to the charts determined for other cycles discussed previously and, therefore, forms a basis of comparison with these cycles for our assumed conditions. The curve on this chart identified as "Expense of Evaporation" is made up of the steam consumption of the quad plus the steam consumption of the "Pauly" minus the evaporation in the "Pauly," which replaces a corresponding amount of exhaust in the system.

Summary

In conclusion we will consider the relative merits of the different cycles discussed in preceding text and submit a graphical comparison of their performance.

Three of the five items making up the steam consumption of the sugar

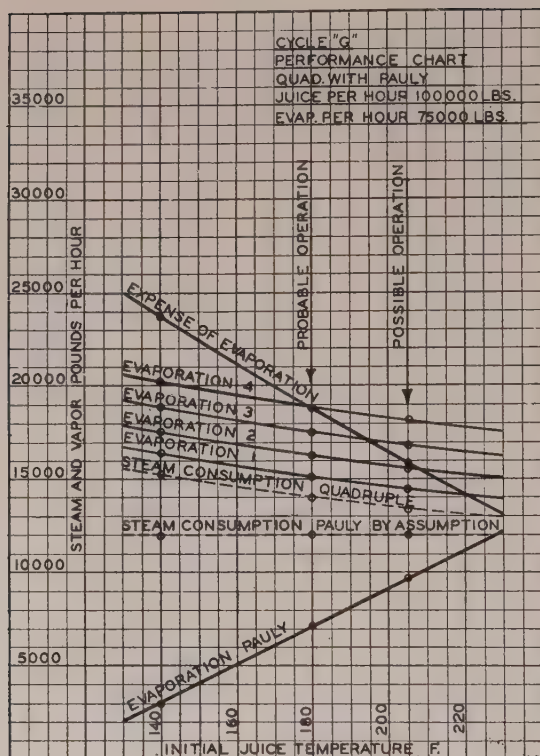


FIG. 27.

factory are assumed as constant and not subject to variation in this discussion. They are:

	Lbs. of Steam F. & A. 212° F.
(1) Power	3,940
(4) Vacuum pans	20,000
(5) Radiation losses	2,500
Fixed total	26,440
Available bagasse steam	54,000
Available for heating and evaporation.....	27,560

The accompanying chart shows the amount of steam required for heating and evaporation according to the various cycles, as compared with the available heat for this purpose.

It is evident that in the case of "A," "B" and "C," it is not possible to operate the factory on bagasse alone when the fibre contents of cane is

less than 10 per cent. Such operation is possible, however, with cycle "F"; "D," if feed is at 198° F.; "E," if feed is at 192° F., and "G," if it is maintained at 203° F. or over. This is, of course, for the specific conditions which we have assumed in making our calculations. It must be clearly borne in mind that deviations from such conditions will alter the case accordingly.

The steam economy is substantially the same for "D" and "E." "D"

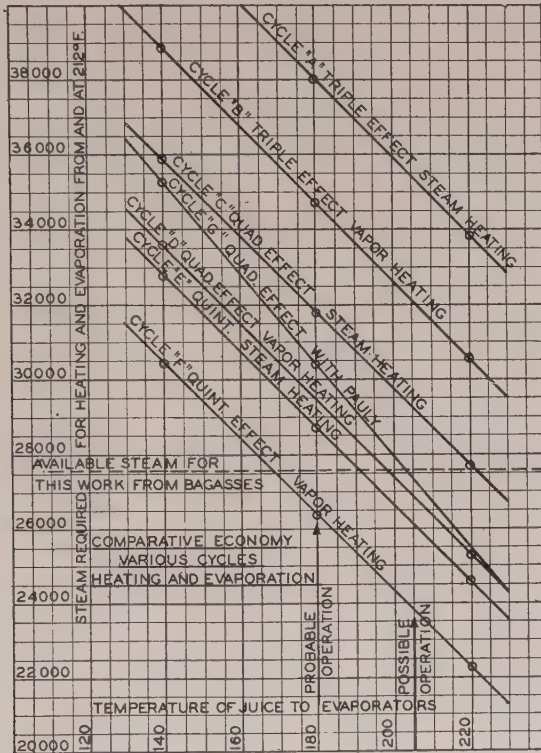


FIG. 28.

represents factory operation with quadruple effect and vapor heater, "E" represents operation with a quintuple effect and steam heater.

The first cost of the "D" equipment is less than the "E" equipment, inasmuch as the heating surfaces are in the ratio of about 4 to 5. The load on the condenser is practically the same in both cases: i.e., 17,850 lbs. per hour for the former and 17,100 lbs. for the latter, so that from this standpoint there is nothing to choose from. From the standpoint of simplicity and uniformity of operation, however, "E" is far preferable to "D." As the juice heating of the cycle "E" is done with exhaust steam, it is entirely separate and independent. As such it is controlled and regu-

lated according to the requirements of the case, permitting a positive temperature control which is very desirable for proper defecation.

In the same way, the quintuple effect is not subject to fluctuations and is easily and simply regulated.

With cycle "D," however, the heating and evaporating operations are intimately inter-related and cannot be separated while the cycle is in operation. As long as conditions are steady and uniform, there is no trouble but if, for some reason, there is a shut down at the mill the cold juice supply to the heater is suddenly stopped, this means that the vapor which normally was condensed by the heater must now be condensed by the second effect which was not designed for this purpose. The result is a readjustment of vacuum and pressure conditions to abnormal proportions. This, of course, temporarily upsets the temperature equilibrium of the quad and after some time the pressure distribution subsides and conforms to the work imposed.

When the mill resumes operation after the temporary shut down, cold juice is suddenly sent through the heater which at once begins to condense vapors at its normal rate, lowering the pressure in the calandria of the second effect. This causes a "flash" in the first effect, speeding up its evaporation and at the same time slowing the operation of the second, third and fourth body. Naturally, the load on the condenser is reduced, which, in turn, increases the vacuum in the last effect. Finally, the operation smooths down and the temperature drops adjust themselves to the normal work of the quadruple for which it was designed.

These fluctuations are undesirable, although they are not quite as serious as they appear, and there are many factories operating satisfactorily today with cycle "D." The operators soon get accustomed to those changes and learn how to compensate for them. Our choice, however, is decidedly for cycle "E."

In our opinion cycle "G" has little to recommend it over cycle "D." It is not only less economical, but it has, to a greater extent, all of the objections of the former. Not only will a stoppage of the mill cause violent fluctuations but, inasmuch as the "Pauly" acts as a pressure-reducing valve between the live and exhaust systems, any variation in the demand on the exhaust lines will be followed by a similar variation in the evaporation therein. In fact, the work of this body is subject to such wide variations that the juice must be maintained at very high levels in order to keep the tubes from becoming dry at slow speed.

One of the bad features of the "Pauly" is that, for a considerable portion of the time, it may run simply as a heater, for which it was not designed. It will be noted by referring to the diagram that for the particular case under consideration, this "Pauly" will consume up to 5000 lbs. of steam per hour before evaporation begins. The conditions described above will exist when its imposed work is within this range and the circulation will be due purely to convection currents and will therefore be very sluggish. In fact, the heating in each tube will be subject to the stream line effect, i.e., the outer portion of the liquor will be heated and the centre core remain relatively cool. This operation results in a very

serious fouling of the heating surface. This, to a great extent, explains the necessity of using high temperature differences to accomplish a relatively small amount of work in a "Pauly" when conditions are unfavorable.

There is another feature worthy of note in connection with these high temperatures. There is a certain quantity of lime salts present in the juices, some of which were in the original juice and some of which were introduced during the defecation by liming. Such lime salts are less soluble when hot than when cold. In the "Pauly" the imposed temperature (227° F.) is 15° F. higher than the defecation, and, therefore, upon passing the defecating point (212° F.) there will be an additional quantity of lime salts precipitated which will adhere immediately to the heating surface. The character and amount will of course depend upon the composition of these lime salts and the quantity in solution.

By referring to the diagram, it will also be seen that when operating normally as indicated the heat transmitted in the "Pauly" is practically equal to that transmitted in the first body of the quad. A very common error is to use insufficient heating surface. If the "Pauly" is to accomplish any economy, it should have practically as much surface as the first body of the quadruple following it. In fact, unless the juice entering is very much hotter than we have assumed, it would be very advantageous to preheat it with a regular multicurrent heater floating on the exhaust line, which would tend to minimize the difficulties referred to above.

We do not favor excessive juice temperatures and one of our objections to the "Pauly" is the high temperature to which the juice is carried. It is a known chemical fact that inversion is more rapid as to temperature increases, especially on low density juices, both of which conditions obtain in this apparatus. It is said that tests have shown these losses to be small, but it is unnecessary to have them at all.

Cycle "F," the quintuple effect with vapor heating, is more economical than any other combination described herein. It has the same objections as cycle "D."

However, it becomes an urgent necessity when there is low fibre content and there is no reason why it should not prove very satisfactory. It has been used in one instance most successfully and was only abandoned because it was too economical, resulting in an excessive accumulation of bagasse under the operating conditions of this particular factory.

Juice Preheating. The performance of all the cycles which have been discussed can be improved by the use of heaters which preheat the juice coming into the evaporator and operate on vapors from the first cell. The saving of steam by this step will be proportional to the possible range of heating.

It is not customary in the Cane Sugar Industry to resort to this method, as it involves another piece of equipment with its necessary upkeep and maintenance. As explained in the text, however, there are certain cycles, such as "G," for instance, where the use of such preheaters would be a decided advantage. There would also be a decided advantage in their installation if the initial temperature of the liquor going to the evaporators is relatively low, such as may be expected if the defecation has been sup-

plemented by bleaching and filtering processes. These steps are necessary in the manufacture of white sugar and are always accompanied with serious radiation losses, resulting in lowering the temperature of the defecated juice.

Other Cycles. Several other cycles are possible, but as yet not in general use in cane sugar factories. A discussion of these cycles with diagrammatic illustration follow. Heat Balances have not been calculated although there are no particular difficulties involved.

By referring to the discussion of the previous cycles the reader can analyze these combinations and decide if they are of value for his particular conditions.

Cycle "H." Cycle "H" is very similar to cycle "D." The essential difference is that instead of the quadruple having a large first cell, from

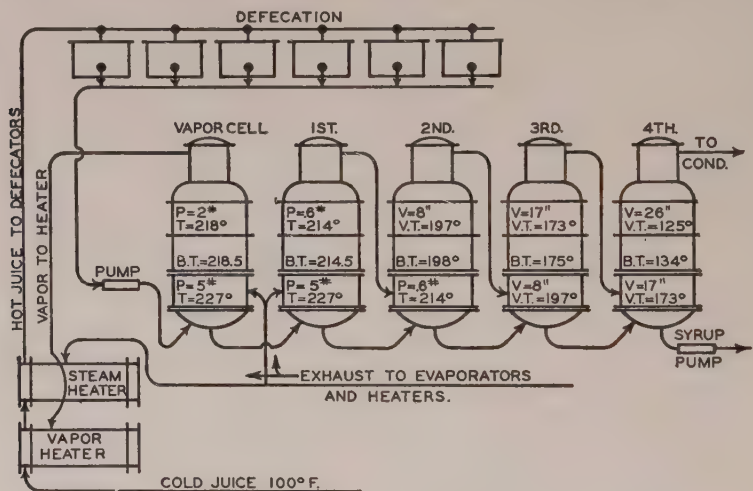


FIG. 29.—Cycle "H"—Quadruple Effect with Vapor Cell.

which vapors are taken out for juice heating, this cell is split up into two units. One of these has its steam belt floating on the exhaust main and its vapor belt operating "dead end" on the heater and the other delivers all of its vapors to the second body of the quadruple. The set is thereby made more flexible and many objections to cycle "D" are overcome. The steam economy is practically the same as "D." Several factories are using this cycle in Cuba.

When the juice supply from the mill ceases on account of some interruption, the boiling liquor in the cell supplying the heater rises in temperature until it reaches an equilibrium with the exhaust main, remains at this temperature until cold juice is again pumped through the heater condensing vapor, when normal operation is again resumed.

Cycle "I." Cycle "I" is simply an elaboration of cycle "H." It has the same quadruple effect as the latter but the vapor for juice heating is

furnished by a pressure double effect instead of a single, as in the former case. Naturally as the steam must go through two stages of evaporation

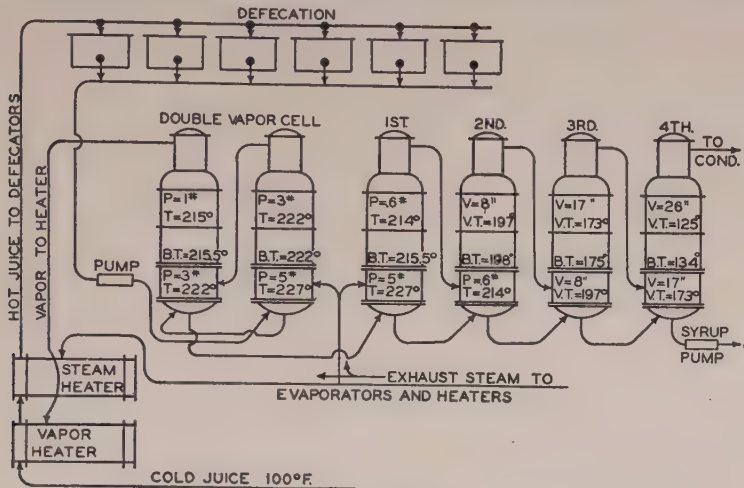


FIG. 30.—Cycle "T"—Quadruple Effect with Double Vapor Cell.

instead of one, within the same temperature fall, each body of this double should have twice as much surface as the single vapor cell in the former

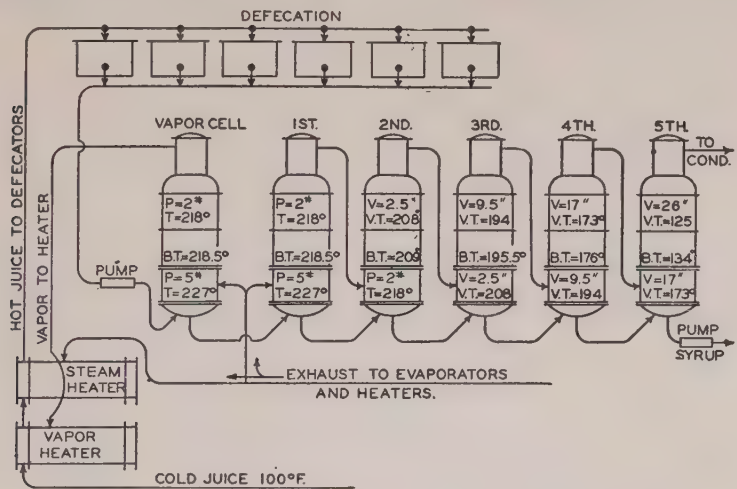


FIG. 31.—Cycle "J"—Quintuple Effect with Vapor Cell.

case. The steam belt of the first effect of the double floats on the exhaust main as before and the vapor space of the second body operates "dead end" on the juice heater.

The economy of this set is considerably better than that of cycle "H" but the first vapor cell is likely to foul very rapidly on account of the fact that the temperature therein is considerably higher than that of defecation. It is therefore essential to provide efficient heating surface and to prepare for frequent cleanings, which can be done by shutting down one cell at a time as the occasion demands.

Cycle "J." Cycle "J" bears the same relation to cycle "F" as cycle "H" does to cycle "D." The large first body of the Quintuple Effect has simply been split up into two parts so as to separate the vapor heating from the Quintuple proper in order to avoid unnecessary complications which disturb the operation. The economy is the same as cycle "F" and the same remarks apply which were made with reference to cycle "H."

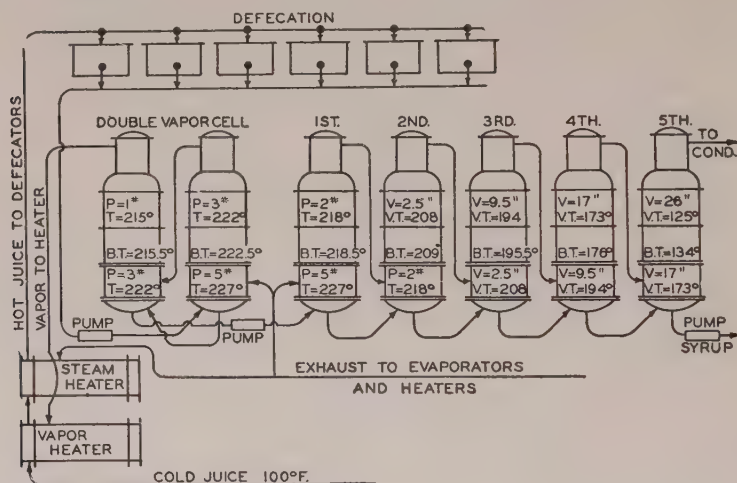


FIG. 32.—Cycle "K"—Quintuple Effect with Double Vapor Cell.

Cycle "K." Cycle "K" is an elaboration of cycle "J." It is similar in all respects to cycle "I" and similar conclusions will apply. Although we do not know of any case where this cycle is in actual use, we see no reason why it is not feasible except as brought out in the discussion of cycle "I."

Cycle "L." Cycle "L" is somewhat involved, being in reality cycle "G" to which has been added a vapor cell. There is at least one large factory using this cycle. It is undoubtedly more economical than cycle "G" but, in our opinion, it is entirely too complicated and unwieldy. We would not recommend its use as better results can be obtained through simpler means.

Cycle "M." One of the oldest and simplest ideas in evaporator work is that of *Vapor Compression* and, in our opinion, it offers large possibilities. We therefore submit cycle "M" representing a possibly successful application to the sugar industry. We understand that three or

four European firms have placed equipment of this type on the market with more or less success. The satisfactory operation of a "Thermo-

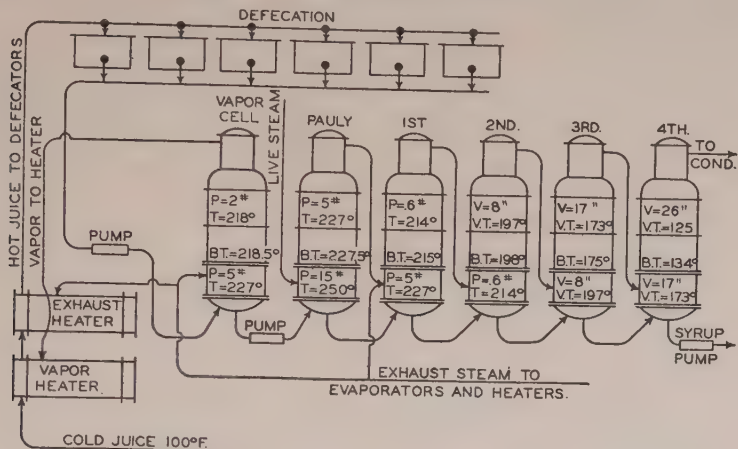


FIG. 33.—Cycle "L"—Quadruple Effect with Vapor Cell and Pauly.

Compressor" depends on many factors. Especially, it is more efficient as the temperature difference between the steam and vapor side of the evaporator to which it is applied is small. It is also more efficient as the

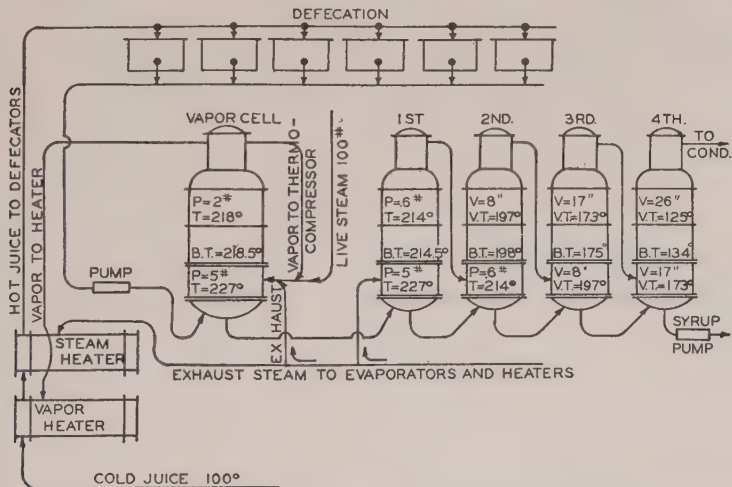


FIG. 34.—Cycle "M"—Quadruple Effect with Vapor Cell and Thermo-Compressor.

specific volume of the vapor being compressed is small; in other words, it is better at or slightly above atmospheric pressure. If the volume of the vapors for a given weight is too great, friction assumes enormous

proportions, as the velocities are very high and the efficiencies, which at best are poor, are greatly lowered.

Under these conditions, we are of the opinion that the logical place for the installation of a "Thermo-Compressor" is on the vapor cell of cycle "H" and the diagram submitted herewith shows such an arrangement.

Another factor to be borne in mind is that the "Thermo-Compressor" cannot be throttled; in other words, it must work at full speed or not at all. This defect can be overcome to a certain extent by installing a battery of small units, which are shut down or turned on one at a time, according to operating conditions.

The compressors are designed to operate against a fixed pressure difference, and if this difference is increased the ratio of compression decreases rapidly. If the pressure relations are further altered the "thermo-compressor" will cease to pull vapors altogether and may even backfire, in which case the live steam will actually flow in the wrong direction and, further, steam from the steam belt will go back to the vapor belt through the throat, actually bypassing exhaust into the vapor space. It is therefore essential to install the Compressor on a body which does not foul and the one which gives the least trouble from this source is the vapor cell. It would be bad practice to install the compressor on a Pauly, which fouls very rapidly and where the pressure difference increases very materially from one cleaning period to the next.

Of course, it must be admitted that as far as we are concerned, the cycle is still in the experimental stage. It promises remarkable results if our small scale data can be successfully translated into practice. The advantages over cycle "H" of course depend entirely upon the "thermo-compressor," which is nothing more or less than a steam injector which compresses vapor by mixing it with live steam. As in cycle "G," the advantage of this arrangement depends upon the proportion of live steam available as compared to exhaust.

A superficial discussion of the thermo-dynamics of the equipment will enable the reader to grasp the principles involved. We will assume that steam at 100 lbs. gauge is available. By referring to the diagram it will be noted that the conditions of operation presume 5 lbs. in the steam belt of the vapor cell and 2 lbs. in the vapor belt. By referring to Steam Tables we find that the total heat of steam for the various conditions is as follows:

For 100 lbs. gauge, or	114.7 lbs.	H = 1,188.8	B.t.u.'s per lb.
" 5 "	" "	19.7 "	H = 1,155.8
" 2 "	" "	16.7 "	H = 1,152.9

Theoretically, therefore, if we wish to convert steam at 2 lbs. gauge into steam at 5 lbs. gauge it is only necessary to add to the former the difference between its total heat and that of the latter or, as seen from the above, 2.9 B.t.u.'s per pound.

When live steam at 100 lbs. gauge is reduced to steam at 5 lbs. gauge there is an excess of heat available equal to the difference of the total heats at these pressures, or 33 B.t.u.'s per pound of live steam, which can be

used for the purpose specified above. Therefore, theoretically, the use of one pound of live steam would enable us to compress $33 \div 2.9$, or 11.4 lbs. of vapor from 2 lbs. gauge to 5 lbs. gauge and, if we could do this, 1 lb. of live steam applied to the vapor cell of the equipment illustrated, would evaporate 11.4 plus 1.0, or 12.4 lbs. of water from the juice. Unfortunately, we are not 100 per cent. efficient, and the best that the author has been able to do under the prescribed conditions, after many years of experimental work on a fairly large scale, was 3 lbs. of water per pound of steam, which is only 24 per cent. efficient.

Nevertheless, even this ratio would result in a very large saving, as can be readily seen. For instance, if we had 9,000 lbs. of live steam avail-

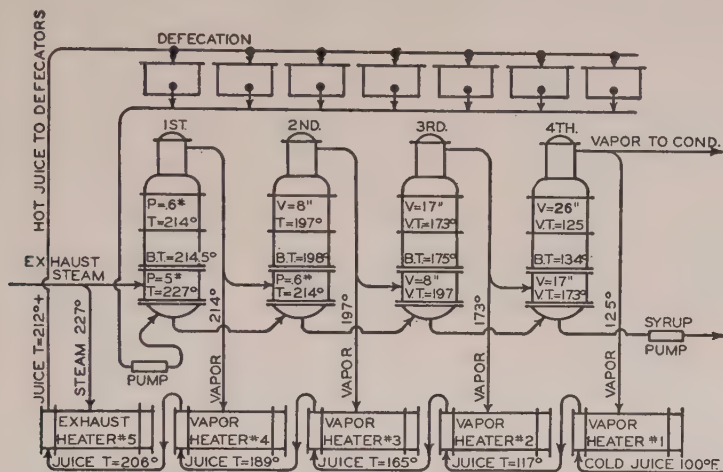


FIG. 35.—Cycle "N"—Multiple Effect with Counter Current Vapor Heating.

able in the factory which we have studied in this text, which is not an impossibility, we would evaporate 18,000 lbs. of water from the juice in the vapor cell without any expenditure as compared with cycle "H" and the net saving over that Cycle would be about 4,000 lbs. of steam per hour. This would bring the total steam consumption of the factory down to about 51,960 lbs. of steam per hour from and at 212° F., if the juice is fed to the evaporators at 180° F., and 49,390 if this temperature is 205° F.

Our opinion is that the "thermo-compressor" offers large possibilities in the field of evaporation, and when its operation is better understood it will have numerous useful applications.

Cycle "N." Cycle "N" represents a method of operation for the use of steam in multiple which is theoretically 100 per cent. efficient, inasmuch as both the evaporation and the heating is done in Multiple Effect. This method may be applied to a system having any number of bodies in the set, although the illustration shows its application to a Quadruple Effect. We know of several installations utilizing this cycle. Practically, the out-

standing objections is the complication involved, the number of heating units with the necessary upkeep and the pumping operations. From the standpoint of heat economy, it is remarkable.

As a matter of interest, we have calculated the heat balance submitted herewith which shows the distribution of B.t.u.'s throughout the system for quadruple operation.

HEAT BALANCE FOR CYCLE "N."

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(1) Steam, 24,600 lbs. @ 227°	$= 24,600 \times 960.7$ B.t.u.'s/lb. =	23,600,000	100,000
Deduct for heating, 100,000 x ($214.5^{\circ} - 180^{\circ} = 34.5^{\circ}$) =		3,450,000	
Available for evaporation.....		20,150,000	
L @ $214^{\circ} = 969.1$ B.t.u.'s/lb. $E1 = \frac{20,150,000}{969.1} =$			20,750
Transferred to No. 2.....			79,250
(2) Vapor ex. No. 1.....		20,150,000	
Liquor flash, 79,250 x ($214.5^{\circ} - 198^{\circ} = 16.5^{\circ}$) =.....		1,308,000	
Available heat		21,458,000	
Deduct heat. No. 4, 100,000 x ($206^{\circ} - 189^{\circ} = 17^{\circ}$) =..		1,700,000	
Available for evaporation.....		19,758,000	
L @ $197^{\circ} = 979.0$ B.t.u.'s/lb. $E2 = \frac{19,758,000}{979.0} =$			20,150
Transferred to No. 3.....			59,100
(3) Vapor ex. No. 2.....		19,758,000	
Liquor flash, 59,100 x ($198^{\circ} - 175^{\circ} = 23^{\circ}$) =.....		1,358,000	
Available heat		21,116,000	
Deduct heat. No. 3, 100,000 x ($189^{\circ} - 165^{\circ} = 24^{\circ}$) =		2,400,000	
Available for evaporation.....		18,716,000	
L @ $173^{\circ} = 994.0$ B.t.u.'s/lb. $E3 = \frac{18,716,000}{994.0} =$			18,850
Transferred to No. 4.....			40,250
(4) Vapor ex. No. 3.....		18,716,000	
Liquor flash, 40,250 x ($175^{\circ} - 134^{\circ} = 41^{\circ}$) =.....		1,650,000	
Available heat		20,366,000	
Deduct heat. No. 2, 100,000 x ($165^{\circ} - 117^{\circ} = 48^{\circ}$) =		4,800,000	
Available for evaporation.....		15,566,000	
L @ $125^{\circ} = 1,021.6$ B.t.u.'s/lb. $E4 = \frac{15,566,000}{1,021.6} =$			15,250
Syrup ex. No. 4.....			25,000
(Cond.) Vapor ex. No. 4.....		15,566,000	
Deduct heat. No. 1, 100,000 x ($117^{\circ} - 100^{\circ} = 17^{\circ}$) =		1,700,000	
Heat to condenser.....		13,855,000	
L @ $125^{\circ} = 1,021.6$ B.t.u.'s/lb. vapor to cond.....			13,560

HEAT BALANCE FOR CYCLE "N" (*Continued*).

Bodies	Specifications	Heat B.t.u.'s	Juice, pounds
(Heat) Ex. heat. No. 5,	$100,000 \times (206^\circ - 212^\circ = 6^\circ) = ..$	600,000	
L @ $227^\circ = 960.7$	B.t.u.'s/lb.		
Steam to heat. No. 5,	$\frac{600,000}{960.7} =$		625
(Sum) Steam consumption evaporator.....			24,600
Steam consumption heater No. 5.....			625
Total			25,225
From and at 212° F. equivalent.....			25,000

It will be noted, by referring to this balance, that the total steam consumption is 25,000 lbs. from and at 212° F. for both heating and evaporation when the feed to the system is 180° F. as before. Further, if this temperature is conserved at 205° F. as mentioned in the foregoing text, there is a further reduction of 2570 lbs. per hour, making the figure 22,450 lbs., which is far below that of any other cycle analyzed previously. The total for the factory is as follows:

Power	3,940
Heating and evaporation.....	25,000
Vacuum pans	20,000
Radiation	2,500
Total	51,440
Available from bagasse alone.....	54,000
Surplus	2,560

This surplus becomes 5130 if the initial temperature of the juice to the evaporators is conserved at 205° F.

This economy is superior even to that of cycle "M," which presupposes the successful application of the Thermo-Compressor.

Greater simplicity but poorer economy is obtainable by omitting some of the heaters. For instance, where we have relatively low vacua, it would undoubtedly pay to use a vapor heater between the last body and the condenser. The heating thus accomplished is a net gain, inasmuch as the heat used would otherwise be all wasted in the condenser. We know of a number of installations where this is done successfully.

The load on the condenser is very materially reduced with counter-current heating, as is shown in the Heat Balance. The amount of steam so condensed is only 13,560 lbs. per hour, which is less than that of any other cycle analyzed.

Other Combinations

There are still further combinations possible in the application of the principles of the Multiple Utilization of Heat in Sugar Factories where extreme economies are essential. However, it is beyond the scope of this book to discuss them fully. We might give as an illustration: the com-

bination of cycles "M" and "N" makes it possible to further reduce the steam consumption of cycle "N" by about 4000 lbs. per hour. These cases, of courses, are special and should be referred to a specialist for study and elaboration, as these analyses require long and careful scrutiny, both as to the thermo-dynamics and their practical application, as the ultimate

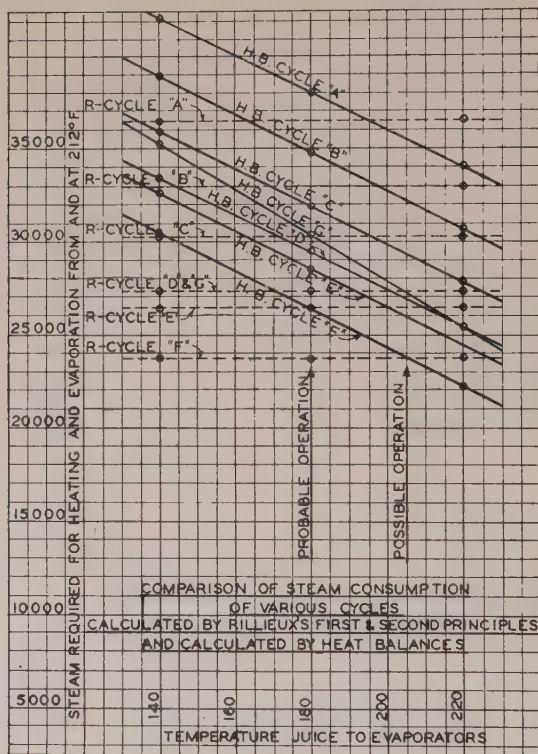


FIG. 36.

purpose is to make a workable and satisfactory installation which will not be unwieldy.

We attach herewith a chart on which is shown a summary of the steam consumption as established by our text and also the steam consumption for the same cycles calculated in accordance with Rillieux's First and Second Principles mentioned above. In the latter, no allowance is made for variations in juice temperatures, latent heat and "flash," so that the result is a very material deviation from the actual conditions.

In conclusion we wish to state that although the above calculations are particularly applicable to the Cuban sugar industry, they can easily be modified to compensate for such variations as may be necessary in other localities.

Chapter 3.

The Heat Balance in the Beet Sugar Factory.

Introductory. The problem encountered in the design of the heating and evaporating stations of the beet sugar factory are similar in a general way to those of the cane sugar factory discussed in the preceding chapter. In certain essentials, however, there are marked differences.

White Sugar. All American beet sugar factories make white sugar directly, instead of making 96 test raws to be subsequently melted and refined in other plants. It is necessary, therefore, to take care of all additional steps involving heat costs, which necessarily must be more expensive than in the manufacture of raw sugar.

No Free Fuel. It must also be remembered that there is no bagasse, the sugar being extracted from the beets by diffusion, leaving the pulp very wet and unfit for fuel. In fact, this pulp is used for other more valuable purposes such as cattle food. Thus all the heat requirements of the factory must be derived from the combustion of outside fuel. Heat economies are therefore pushed much further, so as to keep the costs down to a minimum, and the doctrine of the multiple utilization of heat is carried to a maximum.

Beets are grown in cold climates, often reaching the factory in a frozen condition, from which they have to be thawed. In such climates, heat losses are of much more consequence, and radiation must be given careful attention. Water and juices must be heated through wider ranges, involving ever mounting expense.

Low Vacuum. It is customary in beet sugar factories to carry vacua not exceeding 24 inches. In the first place, the water from the hot wells is used for many purposes, and it is much more desirable hot than at the lower temperatures, which high vacua would yield. This water is used for diffusion, for washing beets, for thawing, for pulp washing, etc., and would have to be heated otherwise. Furthermore, low vacuum is wanted for the operation of pans, as practice has shown that this is the only way to make hard sugar. High vacua produce unmarketable soft-grained products.

High Exhaust Pressure. It has been established by practice that beet juices on account of their chemical characteristics are not seriously injured by moderately high temperatures for short periods of time. Therefore, in order to make up loss of available temperature drop due to carrying low vacua, the pressure on the exhaust mains is raised above that which prevails in the cane sugar industry. Back pressures as high as 18 to 20 lbs. are used, particularly with the new scheme of pressure evaporation. These conditions in turn necessitate higher initial steam pressures in order to provide adequately for the power requirements of the plant.

Based on European Practice. The beet sugar industry in this country has been patterned to a great extent upon European practice, where it was developed. The literature is mostly of European origin, and practically all in the metric system. As pointed out before, in nearly every case, Rillieux's First and Second Principles, with empirical modifications, have been followed in making all heat analyses. These, being at best only crude approximations, will not be used in this book, and all studies given herewith have been carefully and accurately calculated by the heat balance methods which prevail throughout the text.

Specific Case. As in the case of the discussion of the raw cane sugar factory, it is practically impossible to cover all contingencies, and elaboration will therefore be confined to specific problems, basic in character, with the data so arranged that it can be modified to fulfill the particular requirements of those interested. An analysis of the different cycles of operation will be given with a view of bringing out their characteristics, pointing out features of interest where encountered.

These cycles will therefore be based upon cutting 100,000 lbs. of beets per hour, corresponding to the cane factory in the previous chapter. The work will be confined to an elaboration of the heat problems only, assuming that the reader is already familiar with the general processes involved, as well as the details of the filtering and mechanical operations and chemical control.

The amount of diffusion juices from the beets, or "draft" taken from the diffusion batteries, is usually considerably in excess of the weight of such beets, being determined by a number of factors, including the sucrose contents of the beets, the price of sugar, the price of fuel, and other factors depending upon the economics of the enterprise.

Below is given a typical example of the work of a modern beet sugar factory, showing the data necessary to make up the analyses of the various cycles suggested. These data were taken from the actual performance of a factory in operation and therefore represent actual conditions for the operation assumed. The quantities given in this table in the last column represent the weights in percentage of the beets sliced, and give the premises from which the subsequent analyses will be detailed.

Heating Operations. There are many heating operations in the beet sugar factory, and they will be discussed in a general way first.

Contrary to usual practice in this industry, it is assumed that the specific heat of all solutions is unity. Actually, it varies inversely with the density, being less at high concentrations, and more for thin juices, the range being from about 0.9 to 0.65. If the reader desires to make these corrections, they are easily made. The object in making the above assumption is to simplify the calculations in the first place, and to parallel the cane sugar analysis in the second place. The error involved is not serious, as the larger quantities are of low densities, and therefore high specific heats. Such an error will tend to show steam consumptions slightly greater than they should be for heating operations, and slightly smaller for evaporating operations, these two balancing in such a way that the overall calculated

OPERATING CONDITIONS WHITE BEET SUGAR FACTORY.

Nomenclature	Brix.	Analysis		Per Cent. on Beets			Quant. Per Cent. Wt. Beets
		Sucrose	Purity	Solids	Sucrose	Non-Su.	
Beets		18.00	86.0		18.0		100.0
Pulp and Wash Water.					.35		
Diffusion Juice	14.6	12.60	86.5	20.40	17.65	2.75	140.0
Lime, 3% CaO	35.0			3.00			17.5
1st Carb. Juice				25.75	17.65		159.8
1st Cake (55% H ₂ O) ..					.10		13.7
Water to Milk of Lime							14.5
Water to Juice							6.1
Thin Juice	12.9	11.55	89.5	19.60	17.55	2.05	152.0
Thick Juice	64.0	57.20	89.5	19.60	17.55	2.05	30.6
<i>Evaporation Effects</i>							121.4
Thick Juice	64.0	57.20	89.5	19.60	17.55	2.05	30.6
Second Sugar	97.0	93.00	96.0	4.00	3.84	.16	4.1
First Wash	68.0	57.80	85.0	9.23	7.83	1.40	13.6
First Greens to 1st ...	68.0	51.50	75.2	2.55	1.93	.62	3.8
First Melada	92.0	81.20	88.3	35.73	31.50	4.23	38.9
<i>Evaporation 1st Pan</i>							13.2
First Greens to 2nd ...	60.0	51.50	75.2	8.88	6.67	2.21	13.1
Sec. Wash to 2nd Pan.	68.0	49.00	72.0	1.99	1.43	.56	2.9
2nd Sugar	97.0	93.0	96.0	4.00	3.84	.16	4.1
Final Molasses	88.0	51.0	58.0	4.88	2.83	2.05	5.5
Second Melada	94.0	70.0	74.4	10.87	8.10	2.77	11.6
<i>Evaporation 2nd Pan</i>							4.4
1st Grn. and 2nd Wash.	68.0	50.6	74.5	13.42	10.03	3.39	19.8

consumptions will be very slightly greater than they should be, involving a small margin of safety, to which there can be no objection.

It is also assumed that the heaters under consideration will be well designed, with rather fast juice circulation, about 6 to 8 ft. per second. Such heaters will be thoroughly vented and drained, whether they be under pressure or under vacuum, under which conditions it will be easily possible without excessive surfaces to heat juice to within 15° F. of the temperature of steam or vapor under vacuum, and within 10° under pressure. This will be the determining factor in selecting the temperature of steam or vapor to be used in such operations unless otherwise noted.

Evaporators. For the operation of the vacuum evaporators, it is assumed that steam will be available inside the steam chests at 10 lbs. gauge pressure, and vacuum in the condenser at 24 inches referred to a 30-inch barometer, giving temperatures of 240° F. and 141° F., respectively. For the pressure evaporators, it is assumed that exhaust will be available at 17 lbs. gauge pressure in the steam chests, and that the pressure on the vapor side of the last effect will be 1 lb. gauge corresponding to temperatures of 253° F., and 215° F., respectively.

In the distribution of temperature drops between two bodies, practice has indicated that it is approximately correct to assume that there will be an even division of pressures, and the working temperatures are calculated from these. The intermediate pressures and vacua in the discussion that follows have been determined, as per the schedule attached. Undoubtedly,

this is not exactly correct, but so closely approximates the truth that for the purposes of this analysis it may be taken as such.

There is no doubt that as vapors are robbed from the various bodies of any multiple effect in which the concentration of solutions is progressive, from each body to the next, the work is pushed back towards the first body. Even if heating surfaces are proportioned to the transmitted heat, this will not compensate for the change, as the density of the syrup in the cells preceding the last will rise, increasing both the boiling temperature and the viscosity, thus changing the coefficient of heat transmission, and imposing a readjustment of temperature drops to meet these conditions. The extent of such readjustments, however, should be small, and rather than further complicate by additional corrections, it has been assumed that the temperature and pressure distributions of the multiple effects in this discussion will be unaffected by robbing vapors from the various bodies, providing the heating surfaces have been made proportional to the transmitted heat in every case to take care of these vapor heating operations.

SCHEDULE OF VAPOR AND JUICE TEMPERATURES AND LATENT HEATS FOR MULTIPLE EFFECTS.

Basis, Initial Brix, 12.9; Final Brix, 64.0.

System	Nomenclature	Steam Belt 1	Vapor Belt 1	Vapor Belt 2	Vapor Belt 3	Vapor Belt 4	Vapor Belt 5
Triple Effect	Vap. T. ° F.	240.0	222.0	195.0	141.0		
	Boil. T. ° F.		222.5	196.0	148.0		
	B.t.u.						
	L. per pound	952.1	963.9	980.9	1012.6		
Quadruple Effect...	Vap. T. ° F.	240.0	226.0	209.0	185.0	141.0	
	Boil. T. ° F.		226.5	209.5	186.0	148.0	
	B.t.u.						
	L. per pound	952.1	961.3	972.2	986.9	1012.6	
Quintuple Effect ..	Vap. T. ° F.	240.0	230.0	217.0	201.0	179.0	141.0
	Boil. T. ° F.		230.5	217.5	202.0	180.5	148.0
	B.t.u.						
	L. per pound	952.1	958.7	966.5	977.2	990.5	1012.6

Juice Boiler. There is found in most beet sugar factories a so-called *Juice Boiler*, which is a small single effect, operating on live or exhaust steam, and discharging its vapor to the atmosphere. The function of this apparatus is to liberate and expel at least a large part of the ammonia gas always found in beet juices, probably existing as amines which break up on being heated. This operation is most effective as a "scrubbing" process, the juice being sprayed into the vapor space of the boiler thus heating up by direct contact with vapors going out. In this way, on account of the finely divided condition of the liquid, the liberation of ammonia can take place more easily. This cell should operate under first body conditions, or preferably at a vapor pressure slightly in excess of that existing

in the first body which will break up the amines more rapidly on account of higher temperatures. The evaporation from the juice boiler should be very small, as it is a single effect, and therefore very wasteful when considered as an evaporator.

The thin juice, having been deprived of this ammonia gas, will work much better in the multiple effect, the second steam belt of which has a marked tendency to become gas bound otherwise. The evaporation in the juice boiler is relatively so insignificant, that it has not been taken into account in the heat balances. The only loss involved is the vapor going out with the ammonia, and if details brought out above are given due consideration, this need be but a very small quantity.

The Pauly. In the various discussions below, the Pauly will not be taken into consideration. The function of such an apparatus was fully analysed in the chapter on cane sugar, to which the reader is referred. Its value, probably very much exaggerated, is open to question here, and if favorable conditions permit its economical operation, it can be added to any of the cycles outlined.

Vacuum Pans. In order to effect the economies contemplated in the work that follows, it is necessary to assume efficient vacuum pan operation. With particular reference to the use of juice vapors for this work, the pans should be very carefully designed, and of the calandria type, with about $2\frac{3}{4}$ to $2\frac{1}{2}$ sq. ft. of heating surface per cubic foot of melada in the full strike. There must be ample downtakes, and large diameter tubes, 4 to $4\frac{1}{2}$ inches diameter, and 48 to 54 inches long. Such a pan, with well arranged heating surface, thoroughly vented and drained, will operate very well on 4 to 5 lbs. gauge pressure, and even less, particularly at the beginning of the strike. In the analyses of the pressure double effect, full advantage of this possibility has been taken on account of the large amount of vapors available at 215° F., which would have to be wasted if not used in the vacuum pan.

There is another point which it is well to bring out while on the subject of the operation of vacuum pans on juice vapors. The work of the pans is of a variable intermittent nature, while that of the evaporators is continuous and uniform. Success in such operations will therefore depend upon careful timing of the pans, so as to eliminate, as much as possible, the violent changes in heat supply and demand, so that such changes can be accommodated by the evaporating station. This will involve painstaking attention in a small factory, where there are only a few pans, but will become relatively easy in large plants, where there are many pans, the period of whose strikes can be so arranged that the fluctuations will be small.

It is nevertheless true that vapor pan operation will always involve most careful supervision, and will be almost impossible if the operators are allowed to do their work without regard to the operation of the evaporator producing these vapors. The two stations must be under the control of one man, whose duty it will be to properly synchronize the work according to the conditions imposed by the cycle of operation.

As to the actual heat requirements in this station of our typical house

cutting 100,000 lbs. of beets per hour, the schedule below gives this information based upon the original data. The temperature of the feed to the pans is assumed at 195° F. and the boiling temperature in the pan, at 185° F., which represents good practice. The "pan factor" is the ratio of the heat supplied to the pan as compared with its theoretical requirements. This is taken from practise, and is the result of long experience. Some authors use 1.15; the writers, 1.20; others, 1.25. Local conditions, of course, determine this. The heat requirements of the pans have been calculated in pounds of steam from and at 212° F., and also in B.t.u.'s per hour, in fact all quantities are per hour according to the convention assumed earlier.

STEAM CONSUMPTION VACUUM PANS—WHITE BEET SUGAR FACTORY.

(1) First pan (see schedule).

Pan feed per hour.....	52,100 lbs.	
Evaporation	13,200 lbs.	
Feed temperature	195° F.	
Boiling temperature in pan.....	185° F.	
Flash	10° F.	
Heat required, $13,200 \times (L @ 24 \text{ in.} = 1012.6) =$		13,400,000 B.t.u.'s
Deduct flash, $52,100 \times 10 =$		521,000 B.t.u.'s
Total heat =		12,879,000 B.t.u.'s
Pan factor assumed 1.20, heat required.....		15,430,000 B.t.u.'s
Steam required from and at 212 = $\frac{15,430,000}{970.4} =$		15,950 lbs.

(2) Second pan (see schedule).

Pan feed per hour.....	16,000 lbs.	
Evaporation	4,400 lbs.	
Feed temperature	195° F.	
Boiling temperature in pan.....	185° F.	
Flash	10 F.	
Heat required, $4,400 \times (L @ 24 \text{ in.} = 1012.6) =$		4,460,000 B.t.u.'s
Deduct flash, $16,000 \times 10 =$		160,000 B.t.u.'s
Total heat		4,300,000 B.t.u.'s
Pan factor assumed 1.20, heat required.....		5,165,000 B.t.u.'s
Steam required from and at 212 = $\frac{5,165,000}{970.4} =$		5,320 lbs.

Vacuum Triple Effect Cycles.

Preliminary. In making the analyses of the various combinations of evaporators, heaters and pans, experience must be the guide as to what is practical, possible, and advisable, and it would be useless to devote much time to delving into combinations which theoretically appear advantageous, unless such combinations can be put into successful practical use.

This work will therefore be divided into four groups which have been selected from experience as well within the range of commercial requirements. The *first* group will consist of heat analyses for operation with

triple effects, subdivided into three sections: (1) operation with triple effects in which no attempt is made to utilize vapors from any of the bodies of the evaporator; (2) operation with triple effects in which vapor heating is used at every practical opportunity; (3) operation with triple effects, in which not only vapor heating will be incorporated but in which the vacuum pans also will be operated with vapors from the evaporator. The *second* group will comprise similar heat studies, using quadruple effects instead of triples. The *third* group will include similar analyses using quintuple effects. The *fourth* group will be devoted to a study of the possibilities of the newly suggested system of pressure evaporation, with analyses of both the pressure double effect, and the pressure triple effect.

Individual Operations. There are in the beet sugar factory a number of heating operations besides the vacuum pans and the evaporator, the analyses of which must follow a discussion and determination of the latter. Below is given a list of heating operations properly numbered for consideration and study in detail.

HEATING AND EVAPORATING OPERATIONS FOR 100,000 LBS. BEETS PER HOUR. (1) VACUUM TRIPLES

Item	Station	Material	Wt. Lbs. per Hr.	Temperature Range			B.t.u. Required
				From F.	To F.	Rise F.	
1	Diffusion (See calculations separate)	Beets and water	100,000	50	115	65	5,300,000
2	Raw juice	Juice	140,000	113	194	81	11,330,000
3	1st carb.	Juice	159,800	172	194	22	3,520,000
4	2nd carb.	Juice	152,000	185	212	27	4,100,000
5	3rd saturation	Juice	152,000	198	212	14	2,130,000
6	Thin juice	Juice	152,000	200	212	12	1,823,000
7	Ex. evaporators	Thick juice	30,600	140	195	55	1,680,000
8	Syr. and M. S.	Sulph. thick juice	34,700	150	195	45	1,560,000
9	Wash.	Dil. wash.	13,600	115	195	80	1,086,000
10	Greens	High greens	16,000	115	195	80	1,280,000

Item	Station	Material	Wt. Lbs. per Hr.	Temp. In	Bx. Out	Evapo- ration	
11	Evaporators	Juice	152,000	212	12.9	64.0	121,400
12	No. 1 vac. pans	Thick juice, etc.	52,100	195	68.0	92.0	13,200
13	No. 2 vac. pans	Greens, etc.	16,000	195	68.0	94.0	4,400

Item 1—Diffusion. At the diffusion station, the sliced beets are treated in the diffusion battery comprising a number of cells. The general method is to pass water in series through these cells, beginning with cossettes which are nearly exhausted of sugar, and finishing through the lastly filled cell, thus dissolving and removing the sugar from the beets.

The water used for this operation is taken from the condenser hot wells in part and supplemented by extra condensate from the various bodies of the evaporator, from the vapor heaters, and from the caladrias, of the pans when these operate on juice vapors from the evaporator.

Incidentally it might be well to state that the water to feed the boilers

is made up of condensed steam from the various steam-using apparatus, supplemented by part of the condensed vapor above, which is diverted for this purpose.

For our typical factory, the conditions every hour at the diffusion battery would be as follows:

(1) Going to the Diffusion Battery		
(A)	Cossettes (Beets)	100,000 lbs.
(B)	Water to fill cells	102,500 lbs.
(C)	Water to displace juice	132,500 lbs.
Total		335,000 lbs.
(2) Leaving Diffusion Battery		
(A)	Juice Drawn (draft)	140,000 lbs.
(B)	Pulp	93,000 lbs.
(C)	Waste Water	102,000 lbs.
Total		335,000 lbs.

Below is given the heat balance at this station, based on beets at a temperature of 50° F., which has been assumed arbitrarily. It is also assumed that the water from the hot wells of the condensers will arrive at a temperature of 110° F. These figures, more than any of the others, will be determined by local conditions, and it will always be necessary to consider the case on its own merits and conditions in studying the heat requirements of the diffusion battery of any particular plant.

Inasmuch as the quantities in this case are relatively so great, the precedent set forth previously, assuming the specific heat at unity, is not sufficiently precise. In the calculations below, therefore, the specific heats are taken as follows: Water, 1.00; Juice, 0.908; Pulp, 0.900; Cossettes, 0.868. Under these conditions, there are the following:

(2) Heat leaving the battery above 32° F.		
(A)	Juice, 140,000 lbs. @ 115° F. = $140,000 \times (115 - 32 = 83) \times 0.908 = \dots$	10,550,000 B.t.u.'s
(B)	Pulp, 93,000 lbs. @ 110° F. = $93,000 \times (110 - 32 = 78) \times 0.900 = \dots$	6,520,000 B.t.u.'s
(C)	Waste water, 102,000 lbs. @ 110° F. = $102,000 \times (110 - 32 = 78) \times 1.000 = \dots$	7,960,000 B.t.u.'s
Total Heat leaving Battery above 32° F. ...		25,030,000 B.t.u.'s
(1) Heat entering battery above 32° F.		
(A)	Water, 235,000 lbs. @ 110° F. = $235,000 \times (110 - 32 = 78) \times 1.000 = \dots$	18,300,000 B.t.u.'s
(B)	Cossettes, 100,000 lbs. @ 50° F. = $100,000 \times (50 - 32 = 18) \times .868 = \dots$	1,560,000 B.t.u.'s
Total heat entering battery above 32°		19,880,000 B.t.u.'s
Total heat leaving battery above 32°		25,030,000 B.t.u.'s
(See above calculations)		
Deficit to be made up in heaters		5,150,000 B.t.u.'s
Allow 3 per cent. for radiation		150,000 B.t.u.'s
Total heat requirements station 1		5,300,000 B.t.u.'s

This quantity of heat, 5,300,000 B.t.u.'s per hour must be supplied to the "calorizators," or heaters, at the diffusion battery. For this purpose it is advisable to use steam or vapor, providing its temperature is not less than approximately the boiling point at atmospheric pressure, or 212° F., as local conditions impose quick and reliable heating, and the surfaces are limited usually by the arrangement of the cells.

Item 2, Raw Juice. The raw juice, weighing 140,000 lbs. per hour, leaving the diffusion battery above at a temperature of 115° F., arrives at the raw juice heaters at approximately 113° F., having lost 2° on the way, and must be heated at this station to a temperature of 194° F., a range of 81°, involving an expenditure of 11,330,000 B.t.u.'s. This is by far the largest heating operation in the factory and it is customary to split it among two or three heaters, which, when vapor heating is used, are attached to the various bodies of the evaporator, the juice enters the low temperature heater, and leaves the hot heater, the latter being so placed that there remains an operating margin of temperature to make up whatever deficiencies may have developed in the preceding heating units. In each case it is safe to estimate that the juice will be heated to within 15° of the temperature of the vapor used. In the case of the triple effect cycle now under consideration, the range of temperature in each unit will be as follows:

Position of heaters	Cell No. 1	Cell No. 2	Cell No. 3
Vapor temperature	222	195	141
Juice temperature	194-180	180-126	126-113
Heat transferred	1,924,000	7,570,000	1,820,000

Item 3—1st Carbonation. The hot juice at 194° F. goes through the first carbonation, where it is limed and carbonated, and in this process, the temperature drops to 172° F. The weight is increased by this step to 159,800 lbs. The temperature must now be again raised to 194° F., which operation, in the triple effect factory, could be done with vapors from the first body of the evaporator at 222° F., involving a heat expenditure of 3,520,000 B.t.u.'s. It might be possible to split this operation into two parts, with gain in economy, but the complication involved would hardly justify, and for this reason it is done in one operation. It is only in case of very large heating operations that it is justifiable to split them as was done in the first case, for there are always complications involved against which there are inevitable charges to be debited.

Item 4—2nd Carbonation. The hot juice leaving the above station is filtered, and again carbonated, its temperature having dropped to 185° F., and its weight having decreased by removal of sludge in the filters from 159,800 to 152,000 lbs. The temperature must again be raised to 212° F., involving 4,100,000 B.t.u.'s. This step is also carried on with vapors derived from the first cell of the evaporator when vapor heating is used.

Item 5—3rd Saturation. Here, there is another filtration, after which the juice is sulphured, and again heated from its new temperature of 198° F. to 212° F. The weight is practically the same as before, 152,000

lbs. per hour. This also can be done with vapors from the first cell of the evaporator at 222° F., and involves an expenditure of 2,130,000 B.t.u.'s per hour. This heating is done in the juice boiler.

Item 6—Thin Juice. After filtering the sulphured juice subsequent to the heating operation, it is ready for the admission into the evaporator, the clarification now being complete. It has been brought out previously that evaporators are designed to evaporate, and not to heat. It has therefore become universal practise in the beet sugar industry to preheat juice going to the evaporators to as high a temperature as possible using for this purpose vapors from the first cell of the unit. This juice has now been cooled to 200° F. by radiation and otherwise, and it is vapor heated to 212° F. with vapor from No. 1. This step involves 1,823,000 B.t.u.'s, the weight being the same as before, 152,000 lbs. per hour.

It might be interesting to point out at this time that the original juice entered from the diffusion at 115° F., and that after all these steps, it reaches the evaporator at 212° F., a total range of 97° F. Actually, if all the operations are taken into account as detailed above, the range of temperature through which this juice has been raised is 156° F., showing a loss of 59°, or 37.8 per cent. of the heat put in. The contrast with the cane sugar industry is quite marked, for in this case, the juice is heated from 100° F. to 212° F., and reaches the evaporator at 205° F. with modern equipment, showing a loss of only 7° of a total of 112° range, or a loss of 6.25 per cent. The difference, of course, is due to the more elaborate process, and the climatic conditions as noted in the beginning of this discussion.

Item 7—Ex. Evaporators. The thick juice or syrup coming out of the evaporators in a concentrated condition at a temperature of about 140° F. must now be heated to 195° F. for further treatment before going to the pans. The weight is 30,600 lbs. per hour, and this heating is done with vapors from the first cell of the triple if so desired. The heat involved is 1,680,000 B.t.u.'s. Here, also, while the range would suggest multi-stage heating, the quantity is so small, that the complication is not justified, and for this reason, single stage heat is retained.

Item 8—Syrup and Melted Sugar. The second sugar is dissolved in the clarified thick juice from above, changing the weight to 34,700 lbs. per hour. After these operations, the temperature has again dropped to about 150° F., and another heating operation is necessary to raise it again to 195° F. for feeding into the pans, this involves 1,560,000 B.t.u.'s taken from the first cell of the evaporator.

Item 9—Wash Water. The wash water weighs about 13,600 lbs., and must be heated from 115° F., its temperature, to 195° F. to go into the process, involving 1,086,000 B.t.u.'s. This is relatively such a small operation, that it is advisable to do it with steam direct from the boiler, some 33. B. H. P.

Item 10—Greens. The greens weigh about 16,000 lbs. per hour, and must be heated from 115° F. to 195° F., involving 1,280,000 B.t.u.'s. This also is done with steam direct for the same reason as in item 9.

Item 11.—Consideration is reserved until later in the discussion.

Item 12—First Pan. The work of the first pan has been fully discussed earlier, on page 298, to which the reader is referred. The feed to this pan is 52,100 lbs. per hour at a temperature of 195° F. The Brix of the thick juice is 68, and it is concentrated to 92 Brix, involving an evaporation of 13,200 lbs. per hour, involving a heat expense of 15,450,000 B.t.u.'s, which is derived from steam or vapor according to the cycle used.

Item 13—Second Pan. The work of the second pan has also been discussed at the above reference. There is fed into it per hour, 16,000 lbs. of greens, etc., at temperature of 195° F., and a Brix of 68, and the

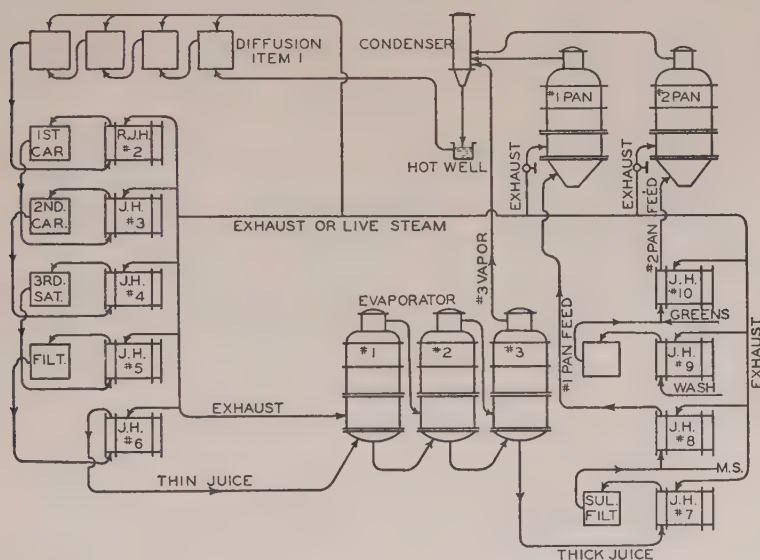


FIG. 1.—Cycle 1-A—Diagrammatic Sketch Triple Steam Heaters—Steam Pans.

concentration goes to 94 Brix, involving an evaporation of 4400 lbs. per hour, necessitating a heat expense of 5,165,000 B.t.u.'s which is derived from steam or vapor according to the cycle.

Item 11—Evaporation. In the group under consideration, the evaporation is done in triple effect, operating under vacuum as explained before. This unit will vary in its work as vapor heaters or vapor pans are attached to it, and each of the three combinations must be studied separately. These, for identification will be called 1-A for the straight triple without vapor heaters or vapor pans; 1-B, with vapor heaters, but no vapor pans; 1-C, with both vapor heaters and vapor pans.

Cycle 1-A. The accompanying heat balance shows the flow of steam, vapor, heat and juice per hour for the straight triple effect, cycle 1-A. This balance has been calculated as all other heat balances in this text, and

hardly needs elaboration in detail. It has been figured to accomplish the imposed work under prescribed conditions, namely, evaporating 121,400 lbs. per hour from 152,000 lbs. of juice per hour at a temperature of 212° F., thus changing its density from an initial of 12.9 Brix, to 64 Brix, the steam supplied to the evaporator being fixed at 10 lbs. gauge, and the vacuum maintained in the condenser at 24 inches, corresponding to a vapor temperature of 141° F. The intermediate conditions of pressure, temperature, boiling points, and latent heats have been given in the beginning of this discussion on page 296, to which the reader is referred.

The attached diagram (Fig. 1) shows the disposition of apparatus for the performance of each function discussed before. In this case, there will be no withdrawal of vapors from the evaporator for any purpose, all of the heating and the vacuum pan operations being done with either live or exhaust steam.

Of course, no one would use such a cycle to-day, and it is given here to show the progression as more bodies are used in the evaporator, and as vapor heating is extended.

The summary sheet for triple effects gives the total heat costs for this as well as other triple effect cycles, and amounts to 95.6 per cent. of the weight of the beets sliced, in steam from and at 212° F. To this figure must be added the heat equivalent of the power used, and the allowance for radiation, both of which are somewhat indeterminate, depending on local conditions. For the purpose of this analysis, it is arbitrarily assumed at 10 per cent. on the weight of the beets for all cycles, which is a very liberal allowance and makes the total for this particular case 105.6 per cent. on the weight of the beets from and at 212° F.

VAPORS AND STEAM FOR PANS AND HEATERS.

VAC. TRIPLE—STEAM HEATERS. CYCLE I-A.

Item	Station	Weight	B.t.u. Req.	Steam	No. 1 Vap.	No. 2 Vap.	No. 3 Vap.
				B.t.u. T = 240	B.t.u. T = 222	B.t.u. T = 195	B.t.u. T = 141
1	Diffusion	B-100,000	5,300,000		None	None	None
2	Raw juice	140,000	11,330,000				
3	1st carb.	159,800	3,520,000				
4	2nd carb.	152,000	4,100,000				
5	3rd sat.	152,000	2,130,000				
6	Thin juice	152,000	1,823,000				
7	Thick juice	30,600	1,680,000				
8	Syr. and M. S.	34,700	1,560,000				
9	Wash.	13,600	1,086,000				
10	Greens	16,000	1,280,000				
Total for heaters.....				33,809,000	33,809,000		
12	No. 1 pan	52,100	15,450,000				
13	No. 2 pan	16,000	5,165,000				
Total for pans.....				20,615,000	20,615,000		

Cycle 1-B. Cycle 1-B is used on triple effect operation as in the preceding case, but vapor heating has been used at every available opportunity, according to the suggestions in the discussion in the beginning of this chapter.

The full heat balance is given below, as well as a diagrammatic sketch (Fig. 2) showing the operations, and a schedule of heating and pan requirements with the dispositions of each.

It must be noted from a study of these data that as more and more vapor heating operations are added to the evaporator, the work is pushed back towards the first body, as pointed out before. It also follows that the load on the condenser also decreases correspondingly.

The steam consumption is shown in the final summary sheet for triple effects, and totals 81.13 per cent. of the weight of the sliced beets in steam from and at 212° F. To this amount must be added the conventional allowance for power and radiation of 10 per cent., making the total 91.13 per cent.

VAPORS AND STEAM FOR PANS AND HEATERS.

VAC. TRIPLE—VAPOR HEATERS. CYCLE 1-B.

Item	Station	Weight	B.t.u. Req.	Steam B.t.u. T = 240	No. 1 Vap. B.t.u. T = 222	No. 2 Vap. B.t.u. T = 195	No. 3 Vap. B.t.u. T = 141
1	Diffusion	B-100,000	5,300,000		5,300,000		
2	Raw juice	140,000	11,330,000		1,940,000 T = 194	7,570,000 T = 180	1,820,000 T = 126
3	1st carb.	159,800	3,520,000		3,520,000 T = 194		
4	2nd carb.	152,000	4,100,000		4,100,000 T = 212		
5	3rd sat.	152,000	2,130,000		2,130,000 T = 212		
6	Thin juice	152,000	1,823,000		1,823,000 T = 212		
7	Syr. thick juice	30,600	1,680,000		1,680,000 T = 195		
8	Syr. and M. S.	34,700	1,560,000		1,560,000 T = 195		
9	Wash.	13,600	1,086,000	1,086,000 T = 195			
10	Greens	16,000	1,280,000	1,280,000 T = 195			
Total for heaters				2,366,000	22,053,000	7,570,000	1,820,000
12	No. 1 pan	52,100	15,450,000				
13	No. 2 pan	16,000	5,165,000				
Total for pans				20,615,000			

THE HEAT BALANCE IN THE BEET SUGAR FACTORY 307

HEAT BALANCE TRIPLE EFFECT—VAPOR HEATING.

Body	Specifications	Heat	Juice
(1)	Steam, 58,600 lbs. @ 240 = $58,600 \times 952.1 =$	55,800,000	152,000
	Deduct for heat, $152,000 \times (222.5 - 212 = 10.5) =$	<u>1,600,000</u>	
	Available for evaporation.....	54,200,000	
	L @ 222 = 963.9, $E_1 = \frac{54,200,000}{963.9} =$		<u>56,300</u>
	Transferred to No. 2		95,700
(2)	Vapor ex. No. 1	54,200,000	
	Add flash, $95,700 \times (222.5 - 196 = 26.5) =$	<u>2,530,000</u>	
	Available heat	56,730,000	
	Deduct for vapor heating (see chart).....	<u>22,953,000</u>	
	Available for evaporation.....	34,677,000	
	L @ 195 = 980.9, $E_2 = \frac{34,677,000}{980.9} =$		<u>35,400</u>
	Transferred to No. 3		60,300
(3)	Vapor ex. No. 2	34,677,000	
	Add flash, $60,300 \times (196 - 148 = 48) =$	<u>2,900,000</u>	
	Available heat	37,577,000	
	Deduct for vapor heating (see chart).....	<u>7,570,000</u>	
	Available for evaporation.....	30,007,000	
	L @ 141 = 1012.6, $E_3 = \frac{30,007,000}{1012.6} =$		<u>29,700</u>
	Thick Juice ex. No. 3		30,600
Condenser			
	Vapor ex. No. 3	30,007,000	
	Deduct for vapor heating (see chart).....	<u>1,820,000</u>	
	Heat to condenser	28,187,000	
	L @ 141 = 1012.6, $E = \frac{28,187,000}{1012.6} =$		<u>27,800</u>
	Steam consumption from and at 212° F. (L @ 212 = 970.4)		
	Heat to No. 1 above = $58,620 \div 970.4 =$		<u>57,500</u>

Cycle 1-C. The third cycle of the triple effect group is the same as the second, except that the vacuum pans are operated on vapor instead of exhaust steam. There results an additional economy in view of the fact that more steam is used in "multiple."

The chart, heat balance and diagram (Fig. 3) follow, giving the same information as in the preceding case.

The total steam consumption for this arrangement is given in the summary sheet, and amounts to 74.33 per cent. of the weight of the beets sliced in pounds of steam from and at 212° F. Making the conventional allowance for power and radiation of 10 per cent., this becomes 84.33 per cent.

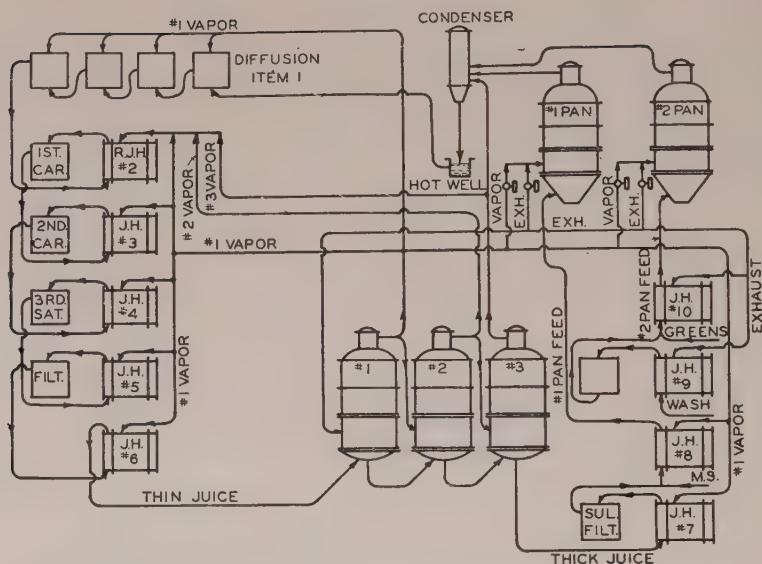


FIG. 3.—Cycle 1-C—Diagrammatic Sketch Triple Effect, Vapor Heaters, Vapor Pans.

VAPORS AND STEAM FOR PANS AND HEATERS.

VAC. TRIPLE—VAP. HEATERS—VAP. PANS. CYCLE 1-C.

Item	Station	Weight	B.t.u. Req.	Steam	No. 1 Vap.	No. 2 Vap.	No. 3 Vap.
				B.t.u. T = 240	B.t.u. T = 222	B.t.u. T = 195	B.t.u. T = 141
1	Diffusion	B-100,000	5,300,000		5,300,000		
2	Raw juice	140,000	11,330,000		1,940,000 T = 194	7,570,000 T = 180	1,820,000 T = 126
3	1st carb.	159,800	3,520,000		3,520,000 T = 194		
4	2nd carb.	152,000	4,100,000		4,100,000 T = 212		
5	3rd sat.	152,000	2,130,000		2,130,000 T = 212		
6	Thin juice	152,000	1,823,000		1,823,000 T = 212		
7	Thick juice	30,600	1,680,000		1,680,000 T = 195		
8	Syr. and M. S.	34,700	1,560,000		1,560,000 T = 195		
9	Wash.	13,600	1,086,000	1,086,000 T = 195			
10	Greens	16,000	1,280,000	1,280,000 T = 195			
12	No. 1. pan	52,100	15,450,000		15,450,000		
13	No. 2 pan	16,000	5,165,000		5,165,000		
Totals				2,366,000	42,718,000	7,570,000	1,820,000

HEAT BALANCE, TRIPLE EFFECT—VAPOR HEATING—VAPOR PANS.

Body	Specifications	Heat	Juice
(1)	Steam, 73,250 lbs. @ 240 = $73,250 \times 952.1 =$	69,750,000	152,000
	Deduct for heat, 152,000 $\times$ (222.5 - 212 = 10.5) =.....	1,600,000	
	Available for evaporation.....	68,150,000	
	L @ 222 = 963.9, E1 = $\frac{68,150,000}{963.9} =$		70,850
	Transferred to No. 2		81,150
(2)	Vapor ex. No. 1	68,150,000	
	Add flash, 81,150 $\times$ (222.5 - 212 = 26.5) =.....	2,150,000	
	Available heat	70,300,000	
	Deduct for heaters and pans (see chart).....	42,718,000	
	Available for evaporation.....	27,582,000	
	L @ 195 = 980.9, E2 = $\frac{27,582,000}{980.9} =$		28,200
	Transferred to No. 3		52,950
(3)	Vapor ex. No. 2	27,582,000	
	Add flash, 52,950 $\times$ (196 - 148 = 48) =.....	2,540,000	
	Available heat	30,122,000	
	Deduct for vapor heaters (see chart).....	7,570,000	
	Available for evaporation.....	22,552,000	
	L @ 141 = 1012.6, E3 = $\frac{22,552,000}{1012.6} =$		22,350
	Thick juice ex. No. 3.....		30,600
	Vapor ex. No. 3	22,552,000	
	Deduct for vapor heaters (see chart).....	1,820,000	
	Heat to condenser	20,732,000	
	L @ 141 = 1012.6, vapor to condenser = $\frac{20,732,000}{1012.6} =$		20,500
	Steam consumption from and at 212° F. (L = 970.4)		
	Heat to No. 1 above = $\frac{69,750,000}{970.4} =$		71,900

Summary of Triple Effect Analyses. The schedule and diagram (Fig. 4) give a summary of the results of the preceding heat studies of the triple effect combinations in the beet sugar factory.

SCHEDULE OF HEAT CONSUMPTION—TRIPLE EFFECTS.

CYCLE I-A—STRAIGHT TRIPLE.

Apparatus	B.t.u.'s Required	Steam, from and at 212	Steam Per Cent. B.
Heaters	33,809,000	34,800	34.8
Pans	20,615,000	21,200	21.20
Evaporators	38,400,000	39,600	39.60
Total	92,824,000	95,600	95.60

CYCLE I-B—TRIPLE AND VAPOR HEATING.

Apparatus	B.t.u.'s Required	Steam, from and at 212°	Steam Per Cent. B.
Heaters	2,366,000	2,430	2.43
Pans	20,615,000	21,200	21.20
Evaporators	55,800,000	57,500	57.50
Total	78,781,000	81,130	81.13

CYCLE I-C—TRIPLE, VAPOR HEATING, VAPOR PANS.

Apparatus	B.t.u.'s Required	Steam, from and at 212°	Steam Per Cent. B.
Heaters	2,366,000	2,430	2.43
Pans			
Evaporators	69,750,000	71,900	71.90
Total	72,116,000	74,330	74.33

The economies to be derived from careful planning are vividly brought out, especially in the diagram Fig. 4. These conclusions are doubly important when it is remembered that the actual cost of total installation for economical operation is probably not more than otherwise, for as the steam consumption is reduced, it must be borne in mind that the requirements of boiler plant capacity are reduced in proportion, which to a great measure should offset the extra cost of other equipment.

As brought out before, the combinations selected are most assuredly not the only ones possible, but fairly represent possible applications easily within the range of practice. If such other combinations are advisable they can be easily analyzed by following the same methods of procedure.

Smaller economies would result by reducing the vapor heating operations, substituting single stage heat for them, or using vapors at higher temperature by robbing the second instead of the third effect, or the first instead of the second.

The heating surfaces of the units under consideration must be determined from the heat balances, by proportioning them to the transmitted amount of heat, and the temperature conditions under which the work is done. This will also be affected by the type of apparatus used, and it is well to make ample allowance for depreciating coefficients, particularly in the heaters, on account of excessive scale accumulation.

Vacuum Quadruple Effect Cycles.

Preliminary. The analysis of the second group of problems, namely, those involving operation with vacuum quadruple effects parallels the preceding discussion. The conditions of operation were given on page 296, to which the reader is referred.

The quantities are the same, and the heating operations are the same except Item 6 which involves heating the juice feed to the evaporator to within 10° F. of the vapor temperature in the first body. This tempera-

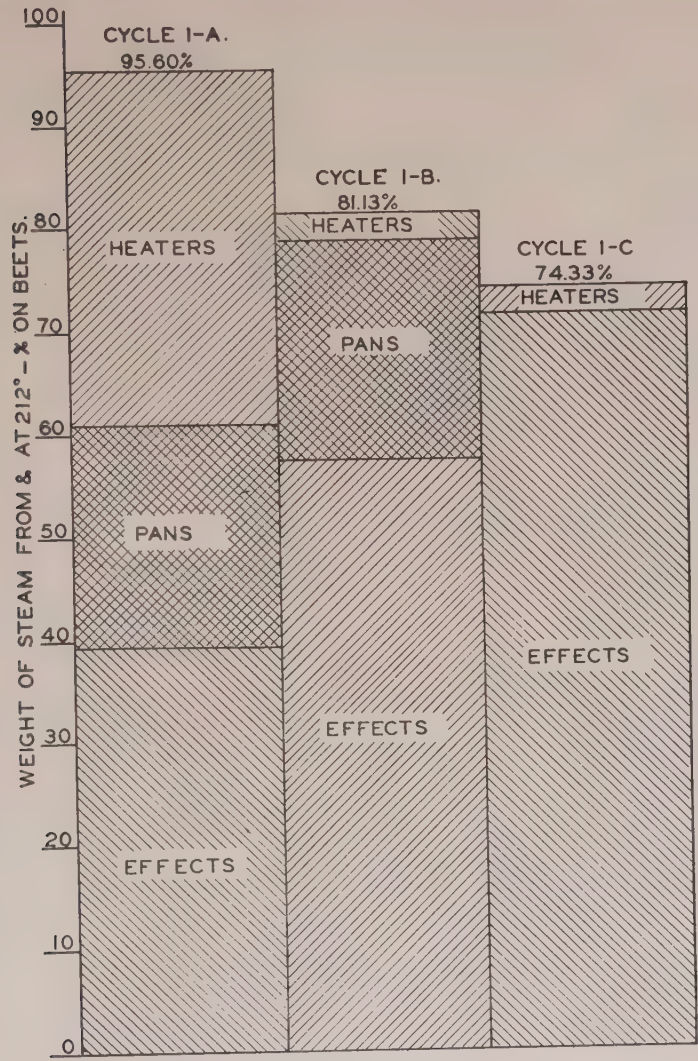


FIG. 4.—Chart Showing Comparative Steam Consumption—Triple Effect Combinations.

ture being different for a quadruple than for a triple, imposes a slight readjustment in the heater in question. Each item is taken up separately below as in the previous case. The accompanying table lists these in detail.

Item 1—Diffusion. At the diffusion battery, the case is identical with the previous, as far as heat requirements are concerned. There must be supplied 5,300,000 B.t.u.'s per hour, and for this purpose, juice vapors

HEATING AND EVAPORATING OPERATIONS FOR 100,000 LBS. BEETS
PER HOUR. (2) VACUUM QUADS.

Item	Station	Material	Wt. Lbs. per Hr.	Temperature Range			B.t.u. Required
				From F.	To F.	Rise F.	
1	Diffusion (See calculations separate)	Beets and water	100,000	50	115	65	5,300,000
2	Raw juice	Juice	140,000	113	194	81	11,330,000
3	1st carb.	Juice	159,800	172	194	22	3,520,000
4	2nd carb.	Juice	152,000	185	212	27	4,100,000
5	3rd saturation	Juice	152,000	198	212	14	2,130,000
6	Thin juice	Juice	152,000	200	216	16	2,435,000
7	Ex. evaporators	Thick juice	30,600	140	195	55	1,680,000
8	Syr. and M. S.	Sulph. thick juice	34,700	150	195	45	1,560,000
9	Wash.	Dil. wash.	13,600	115	195	80	1,086,000
10	Greens	High greens	16,000	115	195	80	1,280,000

Item	Station	Material	Wt. Lbs. per Hr.	Temp.	Bx. In	Bx. Out	Evapo- ration
11	Evaporators	Juice	152,000	216	12.9	64.0	121,400
12	No. 1 vac. pans	Thick juice	52,100	195	68.0	92.0	13,200
13	No. 2 vac. pans	Greens, etc.	16,000	195	68.0	94.0	4,400
Position of Heaters			Cell No. 2	Cell No. 3	Cell No. 4		
Vapor Temperature			209	185	141		
Juice Temperature			194-170	170-126	126-113		
Heat Transferred			3,350,000	6,160,000	1,820,000		

from the second body may be utilized at a temperature of 209° F. if vapors are used. (See temperature distribution chart for each cycle.)

Item 2—Raw Juice. Here, as in the case of triple effect operation, the heating is split up into three heaters drawing heat from the fourth, third, and second effects as shown below. 11,330,000 B.t.u.'s must be supplied to 140,000 lbs. of raw juice per hour, raising its temperature from 113° F. to 194° F., a range of 81° F.

The determining factor in the division of this work is the ability of the heaters to bring the juice temperature up to within 15° of the temperature of the vapor used in them.

Item 3—First Carbonation. At the first carbonation the weight of the juice is 159,800, its temperature must be raised from 172° F. to 194° F. with an expenditure of 3,520,000 B.t.u.'s. If vaporheating is used, this vapor is derived from the first body evaporation at a temperature of 226° F.

Item 4—Second Carbonation. At this station, 152,000 lbs. of juice per hour must be heated from 185° F. to 212° F., which can be done with vapors from the first body as above, at 226° F. The amount of heat is 4,100,000 B.t.u.'s.

Item 5—Third Saturation. At the third saturation the weight is the same as above, 152,000 lbs. of juice per hour must be heated from 198° F. to 212° F., a rise of 14°, involving 2,130,000 B.t.u.'s, which also is supplied with vapors from the first effect if necessary.

Item 6—Thin Juice. The thin juice fed into the evaporator weighs 152,000 lbs. and must be heated from 200° F. to 216° F., a rise of 16°, amounting to 2,435,000 B.t.u.'s, usually derived from the first body vapors whose temperature is 226°.

Item 7—Thick Juice. The thick juice from the last body of the evaporator weighs 30,600 lbs., and is heated from 140° F. to 195° F., a rise of 55°, involving 1,680,000 B.t.u.'s. This is done with vapors from the first effect if desired.

Item 8—Sulphured Thick Juice. This thick juice, with the No. 2 sugar dissolved, weighs 34,700 lbs., and must be heated from 150° F. to 195° F., a rise of 45°, requiring 1,560,000 B.t.u.'s, also taken from the first cell of the evaporator.

Item 9—Dilute Wash. The dilute wash weighs 13,600 lbs., and is heated from 115° F. to 195° F., involving a rise of 80°, requiring an expenditure of 1,086,000 B.t.u.'s. This operation is so small that vapor heating is not justified, and the work is therefore done with direct steam or exhaust.

Item 10—High Greens. The high greens weigh 16,000 lbs. per hour, and are heated from 115° F. to 195° F., a total of 80°, requiring 1,280,000 B.t.u.'s. Direct steam is used for the same reason as above.

Item 11—Reserved for consideration later.

Item 12—No. 1 Pan. This operation is exactly the same as in the case of the triple effect cycles, and involves 15,450,000 B.t.u.'s per hour of vapor or steam as the case may be.

Item 13—No. 2 Pan. The second pan also is the same as in the case of triple effect cycles. There must be supplied 5,165,000 B.t.u.'s per hour of vapor or exhaust.

Item 11—Evaporators. The evaporation is in quadruple effect, beginning with juice, 152,000 lbs. per hour at a temperature of 216° F., and 12.9° Brix; and finishing with 30,600 lbs. of thick juice at 64° Brix, thus having evaporated 121,400 lbs. per hour. The steam requirements vary according to the amounts of vapor taken from the various bodies for heating and vacuum pan operations, as follows in detail.

Cycle 2-A. In cycle 2-A, the evaporation is done in straight quadruple effect, as shown by the heat balance, diagrammatic sketch (Fig. 5) and table of operation. There is no withdrawal of vapors at any point of the evaporator for any purpose, such as heaters and pans, which are all supplied with single stage heat, either live or exhaust steam.

The steam consumptions of the various stations are given in the summary sheets for quadruple effects and consist of an addition of all the heating and vacuum pan operations outlined above, plus the heat requirements of the quadruple effect as shown by the heat balance.

The total is shown to be 85.75 per cent. of the weight of the beets sliced in pounds of steam from and at 212° F. As in the other cases, this amount must be increased to take care of the power requirements, radiation, blow offs, etc., for which is retained the conventional allowance of 10 per cent. of the weight of the beets as before. This brings the total estimated steam consumption of this factory to 95.75 per cent. from and at 212° F.

VAPORS AND STEAM FOR PANS AND HEATERS.

VAC. QUAD—STEAM HEATERS. CYCLE 2-A.

Item	Station	Weight	B.t.u. Req.	Steam B.t.u. T = 240	No. 1 Vap. B.t.u. T = 226	No. 2 Vap. B.t.u. T = 209	No. 3 Vap. B.t.u. T = 185	No. 4 Vap. B.t.u. T = 141
1	Diffusion	B-100,000	5,300,000					
2	Raw juice	140,000	11,330,000					
3	1st carb.	159,800	3,520,000					
4	2nd carb.	152,000	4,100,000					
5	3rd sat.	152,000	2,130,000					
6	Thin juice	152,000	2,435,000					
7	Thick juice	30,600	1,686,000					
8	Syr. and M. S.	34,700	1,560,000					
9	Wash.	13,600	1,086,000					
10	Greens	16,000	1,286,000					
	Total for heaters		34,421,000					
12	No. 1. pan	52,100	15,450,000					
13	No. 2 pan	16,000	5,165,000					
	Total for pans		20,615,000					

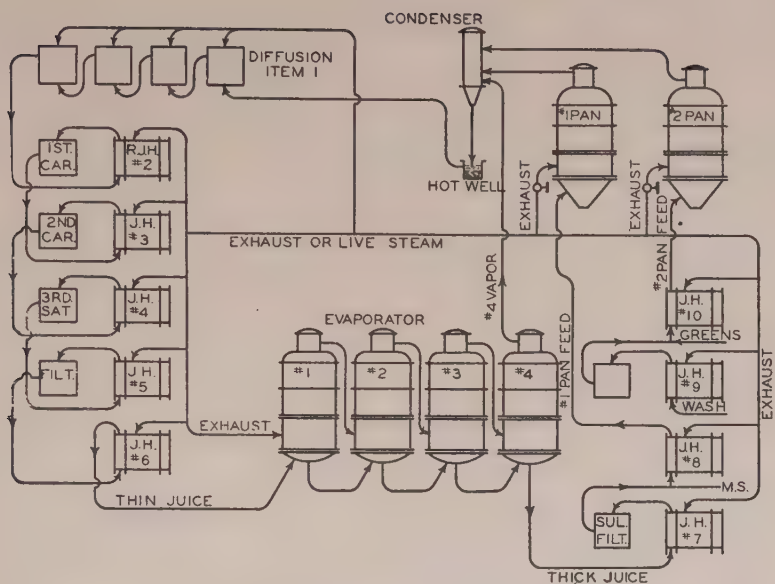


FIG. 5.—Cycle 2-A—Diagrammatic Sketch Quadruple, Steam Heaters, Steam Pans.

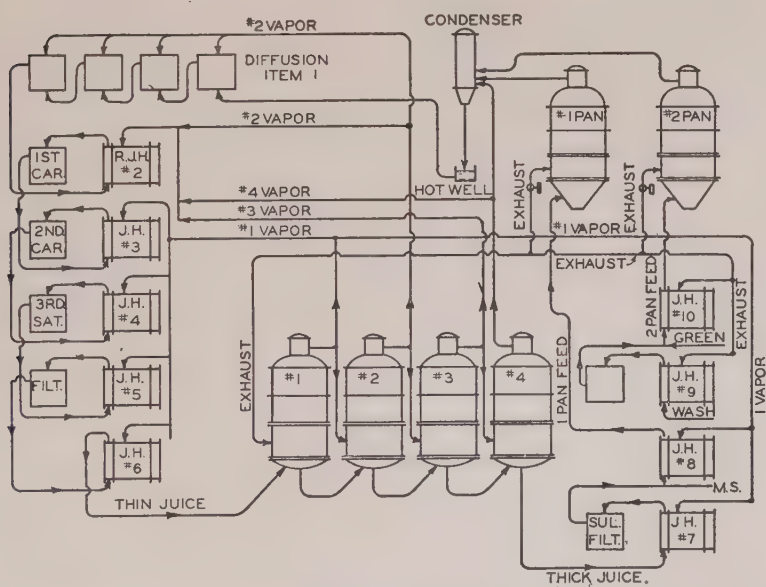


FIG. 6.—Cycle 2-B—Diagrammatic Sketch Quadruple, Vapor Heaters, Steam Pans.

HEAT BALANCE STRAIGHT QUADRUPLE EFFECT.

Body	Specifications	Heat	Juice
(1)	Steam, 29,550 lbs. @ 240 = $29,550 \times 952.1 =$	28,150,000	152,000
	Deduct for heat, $152,000 \times (226.5 - 216 = 10.5) =$	1,595,000	
	Available for evaporation.....	26,555,000	
	L @ 226 = 961.3, $E1 = \frac{26,555,000}{961.3} =$		27,700
	Transferred to No. 2		124,300
(2)	Vapor ex. No. 1	26,555,000	
	Add flash, $124,300 \times (226.5 - 209.5 = 17) =$	2,115,000	
	Available for evaporation.....	28,670,000	
	L @ 209 = 972.2, $E2 = \frac{28,670,000}{972.2} =$		29,500
	Transferred to No. 3		94,800
(3)	Vapor ex. No. 2	28,670,000	
	Add flash, $94,800 \times (209.5 - 186 = 23.5) =$	2,230,000	
	Available for evaporation.....	30,900,000	
	L @ 185 = 986.9, $E3 = \frac{30,900,000}{986.9} =$		31,300
	Transferred to No. 4		63,500
(4)	Vapor ex. No. 3	30,900,000	
	Add flash, $63,500 \times (186 - 148 = 38) =$	2,410,000	
	Available for evaporation	33,310,000	
	L @ 141 = 1012.6, $E4 = \frac{33,310,000}{1,012.6} =$		32,900
	Thick juice ex. No. 4.....		30,600
	Steam consumption from and at 212 (L @ 212 = 970.4)		
	Heat to No. 1 above = $\frac{28,150,000}{970.4} =$		29,000

Cycle 2-B. With cycle 2-B, it is assumed that the various bodies of the quadruple effect are robbed at every available opportunity for supplying heat to the heaters.

The diagram (Fig. 6) indicates the flow of juice and vapors as well as other operations.

The accompanying table shows the distribution of the various operations, showing the stations, the weights of material, the temperatures, and the points from which vapors are robbed at the quadruple. At the bottom is an addition, showing the total amounts taken from each body.

This data is necessary in making up the heat balance to correspond to this work at the evaporator, which balance is shown in detail, indicating the work of the apparatus at all points.

As brought out previously, it is well to mention that the robbing of vapors tends to increase the work of the first bodies of the evaporator, and correspondingly decrease the work of the latter bodies.

The summary sheet shows the heat consumed by various stations of

VAPORS AND STEAM FOR PANS AND HEATERS.

VAC. QUAD—VAPOR HEATERS. CYCLE 2-B.

Item	Station	Weight	B.t.u. Req.	Steam B.t.u. T = 240	No. 1 Vap. B.t.u. T = 226	No. 2 Vap. B.t.u. T = 209	No. 3 Vap. B.t.u. T = 185	No. 4 Vap. B.t.u. T = 141
1	Diffusion	B-100,000	5,300,000			5,300,000		
2	Raw juice	140,000	11,330,000			3,350,000	6,160,000	1,820,000
3	1st carb.	159,800	3,520,000		3,520,000 T = 194	T = 194	T = 160	T = 126
4	2nd carb.	152,000	4,100,000		4,100,000 T = 212			
5	3rd sat.	152,000	2,130,000		2,130,000 T = 212			
6	Thin juice	152,000	2,435,000		2,435,000 T = 216			
7	Thick juice	30,600	1,680,000		1,680,000 T = 195			
8	Syr. and M. S.	34,700	1,560,000		1,560,000 T = 195			
9	Wash.	13,600	1,086,000	1,086,000				
10	Greens	16,000	1,280,000	1,280,000				
	Total for heaters			2,360,000	15,425,000	8,650,000	6,160,000	1,820,000
12	No. 1 pan	52,100	15,450,000					
13	No. 2 pan	16,000	5,165,000					
	Total for pans		20,615,000	20,615,000				

the factory, which totals 70.15 per cent. of the weight of the beets in "steam from and at 212° F." To this must be added the conventional allowance of 10 per cent., making the total steam consumption 80.15 per cent. of the weight of the beets chipped in steam from and at 212° F.

HEAT BALANCE QUADRUPLE EFFECT VAPOR HEATING.

Body	Specifications	Heat	Juice
(1)	Steam, 48,300 lbs. @ 240 = $48,300 \times 952.1 =$	46,000,000	152,000
	Deduct for heat, $152,000 \times (226.5 - 216 = 10.5) =$	1,595,000	
	Available for evaporation.....	44,405,000	
	L @ 226 = 961.3, $E_1 = \frac{44,405,000}{961.3} =$		46,200
	Transferred to No. 2		105,800
(2)	Vapor ex. No. 1	44,405,000	
	Add flash, $105,800 \times (226.5 - 209.5 = 17) =$	1,800,000	
	Available heat	46,205,000	
	Deduct for vapor heating (see chart).....	15,425,000	
	Available for evaporation.....	30,780,000	
	L @ 209 = 972.2, $E_2 = \frac{30,780,000}{972.2} =$		31,700
	Transferred to No. 3		74,100
(3)	Vapor ex. No. 2	30,780,000	
	Add flash, $74,100 \times (209.5 - 186 = 23.5) =$	1,740,000	
	Available heat	32,520,000	
	Deduct for vapor heating (see chart)	8,650,000	
	Available for evaporation.....	23,870,000	
	L @ 185 = 986.9, $E_3 = \frac{23,870,000}{986.9} =$		24,200
	Transferred to No. 4		49,900
(4)	Vapor ex. No. 3	23,870,000	
	Add flash, $49,900 \times (186 - 148 = 38) =$	1,900,000	
	Available heat	25,770,000	
	Deduct for vapor heating (see chart)	61,160,000	
	Available for evaporation.....	19,610,000	
	L @ 141 = 1012.6, $E_4 = \frac{19,610,000}{1,012.6} =$		19,300
	Thick juice ex. No. 4.....		30,600
Condenser	Vapor ex. No. 4	19,610,000	
	Deduct for vapor heaters (see chart).....	1,820,000	
	Vapor heat to condenser.....	17,690,000	
	L @ 141 = 1012.6, Vapor = $\frac{17,690,000}{1012.6} =$		17,450
	Steam consumption from and at 212° F. (L @ 212 = 970.4)		
	Heat to No. 1 above = $\frac{46,000,000}{970.4} =$		47,500

Cycle 2-C. Cycle 2-C is an elaboration of cycle 2-B, in which, not only are the heating operations done with vapor wherever possible, but in addition, the vacuum pans are operated on vapors.

Fig. 7 shows the combinations as in the previous case, indicating the relations between the various apparatus, as well as their functions.

The heat balance shows the complete operation of the quadruple effect to accomplish the work contemplated, with all quantities given in pounds per hour, according to the convention prevailing throughout the discussion.

The steam consumption is shown in the final summary sheet for the quadruple effect operation. It is 64.1 per cent. of the weight of the beets

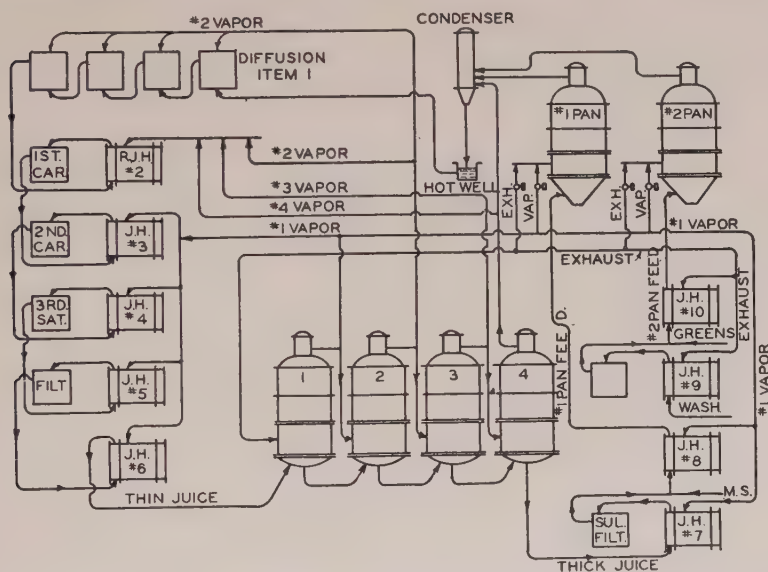


FIG. 7.—Cycle 2-C—Diagrammatic Sketch Quadruple, Vapor Heaters, Vapor Pans.

sliced, in steam from and at 212° F. Adding the usual allowance for power, radiation, etc., this becomes 74.1 per cent.

Summary of Quadruple Effect Analyses. The accompanying summary sheet shows in brief the conclusions derived from the study of the possible and desirable operation of quadruple effects in the beet sugar factory under consideration.

Fig. 8 shows the same data graphically, bringing out very distinctly the savings of fuel possible by the use of the principles of multiple utilization of heat, so little understood by many engaged in this industry.

As in the previous case, the schemes utilized represent what in the opinion of the author would be good practice. They are not the only ones possible by any means, for space does not permit the consideration of them all. Local conditions will impose modifications. After all, the pur-

VAPORS AND STEAM FOR PANS AND HEATERS.

VAC. QUAD—VAP. HEATERS—VAP. PANS, CYCLE 2-C.

Item	Station	Weight	B.t.u. Req.	Steam B.t.u. T = 240	No. 1 Vap. B.t.u. T = 226	No. 2 Vap. B.t.u. T = 209	No. 3 Vap. B.t.u. T = 185	No. 4 Vap. B.t.u. T = 141
1	Diffusion	B-100,000	5,300,000			5,300,000		
2	Raw juice	140,000	11,330,000			3,350,000	6,160,000	1,820,000
3	1st carb.	159,800	3,520,000		3,520,000 T = 194	T = 194	T = 160	T = 126
4	2nd carb.	152,000	4,100,000		4,100,000 T = 212			
5	3rd sat.	152,000	2,130,000		2,130,000 T = 212			
6	Thin juice	152,000	2,435,000		2,435,000 T = 216			
7	Thick juice	30,600	1,680,000		1,680,000 T = 195			
8	Syr. and M. S.	34,700	1,560,000		1,560,000 T = 195			
9	Wash.	13,600	1,086,000	1,086,000				
10	Greens	16,000	1,280,000	1,280,000				
	Total for heaters			2,360,000	15,425,000	8,650,000	6,160,000	1,820,000
12	No. 1 pan	52,100	15,450,000		15,450,000			
13	No. 2 pan	16,000	5,165,000		5,165,000			
	Total for pans		20,615,000		20,615,000			

CYCLE 2-C.

HEAT BALANCE, QUADRUPLE, VAPOR HEATERS, VAPOR PANS.

Body	Specifications	Heat	Juice
(1)	Steam, 64,800 lbs. @ 240 = $64,800 \times 952.1 =$	61,700,000	152,000
	Deduct for heaters, $152,000 \times (226.5 - 216 = 10.5) =$..	1,595,000	
	Available for evaporation.....	60,105,000	
	L @ 226 = 961.3, $E1 = \frac{60,105,000}{961.3} =$		62,500
	Transferred to No. 2		89,500
(2)	Vapor ex. No. 1	60,105,000	
	Add flash, $89,500 \times (226.5 - 209.5 = 17) =$	1,520,000	
	Available heat	61,625,000	
	Deduct for heaters and pans (see chart).....	36,040,000	
	Available for evaporation.....	25,585,000	
	L @ 209 = 972.2, $E2 = \frac{25,585,000}{972.2} =$		26,300
	Transferred to No. 3		63,200
(3)	Vapor ex. No. 2	25,585,000	
	Add flash, $63,200 \times (209.5 - 186 = 23.5) =$	1,485,000	
	Available heat	27,070,000	
	Deduct for vapor heating (see chart).....	8,650,000	
	Available for evaporation.....	18,420,000	
	L @ 185 = 986.9, $E3 = \frac{18,420,000}{986.9} =$		18,650
	Transferred to No. 4		44,550
(4)	Vapor ex. No. 3	18,420,000	
	Add flash, $44,550 \times (186 - 148 = 38) =$	1,690,000	
	Available heat	20,110,000	
	Deduct for vapor heating (see chart).....	6,160,000	
	Available for evaporation.....	13,950,000	
	L @ 141 = 1012.6, $E4 = \frac{13,950,000}{1012.6} =$		13,800
	Thick juice ex. No. 4.....		30,750
Condenser			
	Vapor ex. No. 4	13,950,000	
	Deduct for vapor heaters (see chart).....	1,820,000	
	Heat to condenser	12,130,000	
	L @ 141 = 1012.6, vapor to condenser = $\frac{12,130,000}{1012.6} =$		12,000
	Steam consumption from and at 212° F. (L @ 212 = 970.)		
	Heat to No. 1 above = $\frac{61,700,000}{970.4} =$		63,700

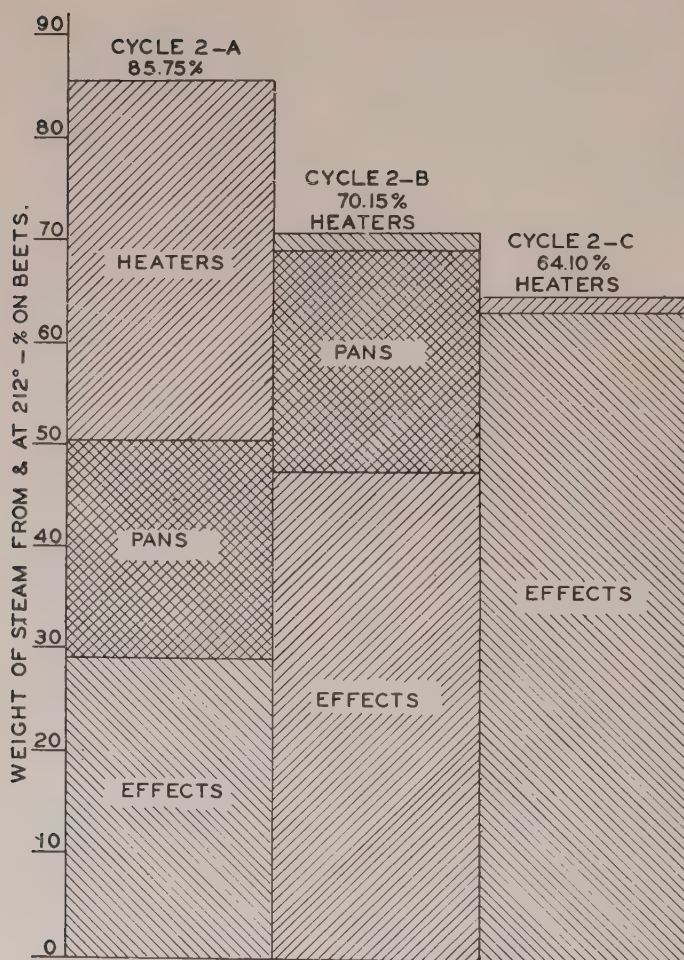


FIG. 8.—Chart Showing Comparative Steam Consumption—Quadruple Effect Combinations.

SUMMARY SHEET—SCHEDULE OF HEAT CONSUMPTION—QUADRUPL E EFFECTS.

CYCLE 2-A—STRAIGHT QUADRUPL E.

Apparatus	B.t.u.'s Required	Steam, from and at 212° F.	Steam Per Cent. B.
Heaters	34,421,000	35,500	35.50
Pans	20,615,000	21,250	21.25
Evaporators	28,150,000	29,000	29.00
Total	83,186,000	85,750	85.75

CYCLE 2-B--QUADRUPL E AND VAPOR HEATING.

Apparatus	B.t.u.'s Required	Steam, from and at 212° F.	Steam Per Cent. B.
Heaters	1,366,000	1,405	1.40
Pans	20,615,000	21,250	21.25
Evaporators	46,000,000	47,500	47.50
Total	68,081,000	70,155	70.15

CYCLE 2-C--QUADRUPL E, VAPOR HEATING, VAPOR PANS.

Apparatus	B.t.u.'s Required	Steam, from and at 212° F.	Steam Per Cent. B.
Heaters	1,366,000	1,405	1.40
Pans			
Evaporators	61,700,000	63,700	63.70
Total	63,066,000	64,105	64.10

pose of this book is to attack only the specific problem set forth in the beginning of this chapter, thus giving a general idea of the proper method of procedure, applicable to any case.

Vacuum Quintuple Effect Cycles.

Preliminary. The third group of the series includes combinations for the beet sugar industry involving the use of five bodies to the evaporator, or a quintuple effect. This case is similar in all respects to the preceding

HEATING AND EVAPORATING OPERATIONS FOR 100,000# BEETS PER HOUR. (3) VACUUM QUINT.

Item	Station	Material	Wt. Lbs. per Hr.	Temperature Range			B.t.u. Required
				From F.	To F.	Rise F.	
1	Diffusion (See calculations separate)	Beets and water	100,000	50	115	65	5,300,000
2	Raw juice	Juice	140,000	113	194	81	11,330,000
3	1st carb.	Juice	159,800	172	194	22	3,520,000
4	2nd carb.	Juice	152,000	185	212	27	4,100,000
5	3rd saturation	Juice	152,000	198	212	14	2,130,000
6	Thin juice	Juice	152,000	200	220	20	3,040,000
7	Ex. evaporators	Thick juice	30,600	140	195	55	1,680,000
8	Syr. and M. S.	Sulph. thick juice	34,700	150	195	45	1,560,000
9	Wash.	Dil. wash.	13,600	115	195	80	1,086,000
10	Greens	High greens	16,000	115	195	80	1,280,000
Item	Station	Material	Wt. Lbs. per Hr.	Temp.	Bx. In	Bx. Out lbs.	Evapo- ration per Hr.
11	Evaporators	Juice	152,000	220	12.9	64.0	121,400
12	No. 1 vac. pans	Thick juice	52,100	195	68.0	92.0	13,200
13	No. 2 vac. pans	Greens, etc.	16,000	195	68.0	94.0	4,400

one, the specific analyses being confined to the same hypothetical factory discussed earlier in the chapter.

The arrangement is designed to take advantage of every opportunity of saving heat in accordance with the basic ideas previously set forth. The distribution of temperatures, pressures, vacua, etc., are outlined on page 296. The quantities are the same as before, as also the work done, and the heating operations with the exception of item 6, in which the juice is heated to a higher temperature in view of the fact that the boiling point in the first body of the evaporator is higher than in the preceding cases, where triples and quadruples were under discussion. The accompanying table gives each operation in detail. A discussion of these follows:

Item 1—Diffusion. The heat requirements of the diffusion battery are 5,300,000 B.t.u.'s per hour. For the straight quintuple effect, this is derived from live or exhaust steam. Where vapor heating is used, this heat comes from vapors leaving the second body of the evaporator at a temperature of 217° F. Theoretically, cheaper heat might be used, and it has been well established that it is inadvisable to use vapor below atmospheric pressure, and that is the reason for the selection. The range of temperature is of course relatively low, being from 50° F. to 115° F., and under ordinary conditions, this work could be done even with vapors going to the condenser. Practice has shown, however, that this cannot be done successfully.

Item 2—Raw Juice. In the heating of raw juice, there is a very large amount of heat required, and through a considerable range, which gives a very good opportunity to use cheap vapors. The amount involved is 11,330,000 B.t.u.'s, and the range from 113° F. to 194° F. In the case of the straight quintuple, this is done with live or exhaust steam, as are all other heating operations. When vapor heating is used, this is split up into several sections using first the vapors going from the last body to the condenser, bringing the temperature to within 15° of that of the vapors; next, vapors from the fourth body are used as far as possible, and so on, until the required temperature is reached, as outlined in the schedule below:

Position Heat.	Cell No. 2	Cell No. 3	Cell No. 4	Cell No. 5
Vapor Temperature	217	201	179	141
Juice Temperature	186-194	164-186	126-164	113-126
Heat Transfers B.t.u.'s	1,110,000	3,080,000	5,320,000	1,820,000

Item 3—First Carbonation. The heating station at the first carbonation must supply 3,520,000 B.t.u.'s from a temperature of 172° F. to 194° F. Part of this could be done with vapors from the third body at 201° , supplemented with vapors from the second cell at 217° . This, however, would result in two very small operations which do not justify the complication. The whole operation is therefore done with vapors from the second body at 217° .

Item 4—Second Carbonation. The second carbonation requires 4,100,000 B.t.u.'s, and the range of heating is from 185° F. to 212° F. In this case also, the theoretical possibilities will not pay their way in practice,

and the heating is most advisable with vapors from the first body at 230° F., and is so calculated.

Other Items. Items 5, 6, 7, 8 are all done with vapors from the first body at 230° F. for the same reason as in Item 4.

Item 9—Wash. **Item 10—Greens.** The heating of the wash water and the greens are very small operations, and as such do not justify the complications involved in vapor heating. This work is done with live or exhaust steam.

Item 11. Reserved for consideration later.

Item 12—No. 1 Pan. The operation of the first pan is exactly the same as in the case of the other cycles analyzed previously. This involves

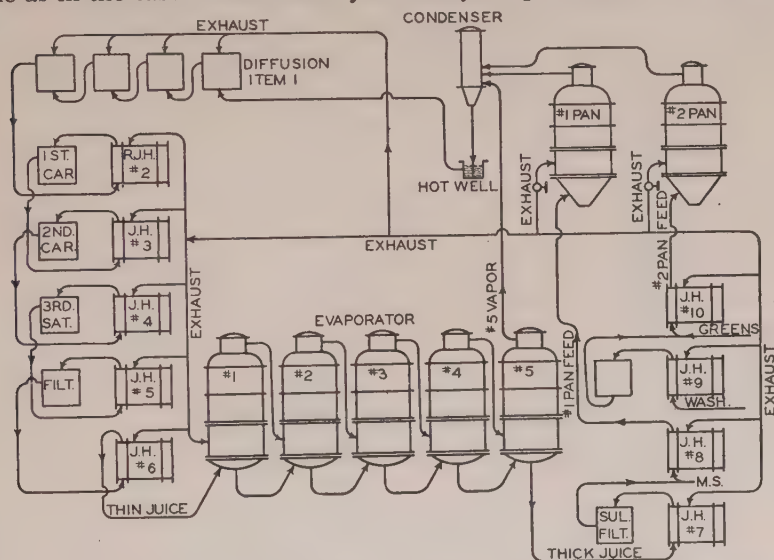


FIG. 9.—Cycle 3-A—Diagrammatic Sketch Quintuple—Steam Heating—Steam Pans. the expenditure of 15,450,000 B.t.u.'s, which is made up of either exhaust steam or vapor from the first effect as the case may be.

Item 13—No. 2 Pan. The same remarks apply to the second pan. The heat requirements are 5,165,000 B.t.u.'s, made up of vapor or exhaust.

Item 11—Evaporators. The evaporation is in quintuple effect, beginning with 152,000 lbs. of thin juice per hour at a temperature of 220° F., and a density of 12.9 Brix, and finishing with 30,600 lbs. per hour of thick juice at 64° Brix, thus evaporating 121,400 lbs. of water per hour from the said thin juice. The amount of steam required for the operation of the apparatus varies according to the quantities of vapors robbed from its various bodies for heaters and pans. This will be detailed below.

Cycle "3-A." Operation according to cycle "3-A" is based on a straight quintuple effect, that is, without the withdrawals of vapors from any of the bodies of the evaporator for any purpose. All of the heating is done with

VAPORS AND STEAM FOR PANS AND HEATERS.

VAC. QUINT.—STEAM HEATERS. CYCLE "3-A."

Item	Station	Weight	B.t.u. Req.	Steam B.t.u. T = 240	No. 1 Vap. B.t.u. T = 230	No. 2 Vap. B.t.u. T = 217	No. 3 Vap. B.t.u. T = 201	No. 4 Vap. B.t.u. T = 179	No. 5 Vap. B.t.u. T = 141
1	Diffusion	B-100,000	5,300,000						
2	Raw juice	140,000	11,330,000						
3	1st carb.	159,800	3,520,000						
4	2nd carb.	152,000	4,100,000						
5	3rd sat.	152,000	2,130,000						
6	Thin juice	152,000	3,040,000						
7	Thick juice	30,600	1,680,000						
8	Syr. and M. S.	34,700	1,560,000						
9	Wash.	13,600	1,086,000						
10	Greens	16,000	1,280,000						
Total for heaters				35,026,000					
12	No. 1 pan	52,100	15,450,000						
13	No. 2 pan	16,000	5,165,000						
Total for pans				20,615,000					

either live or exhaust steam, as well as the operation of the pans. In other words, only Rillieux's first principle is used, omitting the second altogether.

The table accompanying shows the distribution of heat for various operations. Fig. 9 also shows an outline of the various apparatus and their interrelations. The heat balance shows the work of the quintuple effect, with the flow of heat and juice fully detailed, as well as the progressive flow of heat during the work. All the quantities are shown in pounds per hour as before.

The total consumptions are shown in the summary page for quintuple effects. It is shown that the total steam requirements of the factory are equivalent to 79.95 per cent. of the weight of the beets sliced in pounds "from and at 212°." The usual margin of 10 per cent. of the weight of the beets must be added for various contingencies, including radiation, power, etc., which brings the figure to 89.95 per cent.

CYCLE "3-A."—HEAT BALANCE STRAIGHT QUINTUPLE EFFECT.

Body	Specifications	Heat	Juice
(1) Steam, 23,000 lbs. @ 240 = $23,000 \times 952.1 =$		21,950,000	152,000
Deduct for heat. $152,000 \times (230.5 - 220 = 10.5) =$		1,597,000	
Available for evaporation.....		20,353,000	
L @ 230 = 958.7 , $E_1 = \frac{20,353,000}{958.7} =$			21,250
Transferred to No. 2			130,750
(2) Vapors ex. No. 1		20,353,000	
Add flash, $130,750 \times (239.5 - 217.5 = 13) =$		1,700,000	
Available for evaporation.....		22,053,000	
L @ 217 = 966.5 , $E_2 = \frac{22,053,000}{966.5} =$			22,850
Transferred to No. 3			107,900
(3) Vapors ex. No. 2		22,053,000	
Add flash, $107,900 \times (217.5 - 202 = 15.5) =$		1,680,000	
Available for evaporation.....		23,732,000	
L @ 201 = 977.2 , $E_3 = \frac{23,732,000}{977.2} =$			24,300
Transferred to No. 4			83,600
(4) Vapors ex. No. 3		23,732,000	
Add flash, $83,600 \times (202 - 180.5 = 21.5) =$		1,800,000	
Available for evaporation.....		25,532,000	
L @ 179 = 990.5 , $E_4 = \frac{25,532,000}{990.5} =$			25,800
Transferred to No. 5			57,800
(5) Vapors ex. No. 4		25,532,000	
Add flash, $57,800 \times (180.5 - 148 = 32.5) =$		1,880,000	
Available for evaporation.....		27,412,000	
L @ 141 = 1012.6 , $E_5 = \frac{27,412,000}{1012.6} =$			27,150
Thick juice ex. No. 5.....			30,650

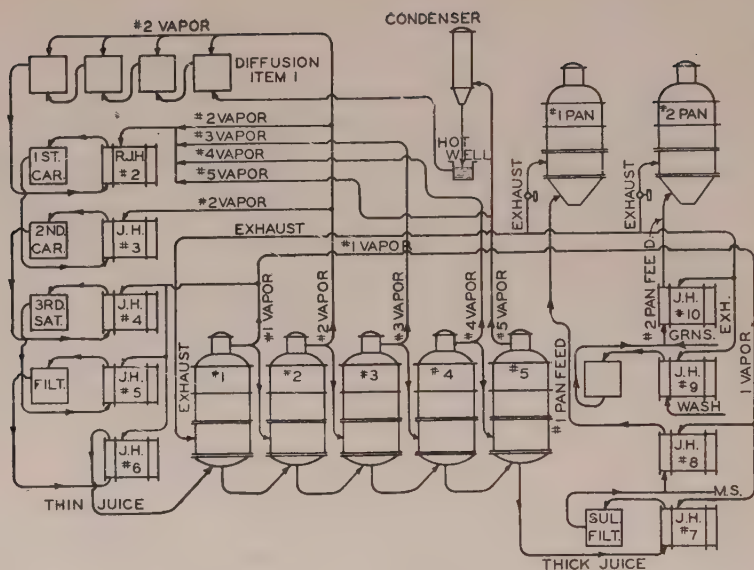


FIG. 10.—Cycle "3-B"—Diagrammatic Sketch Quintuple—Vapor Heating—Steam Pans.

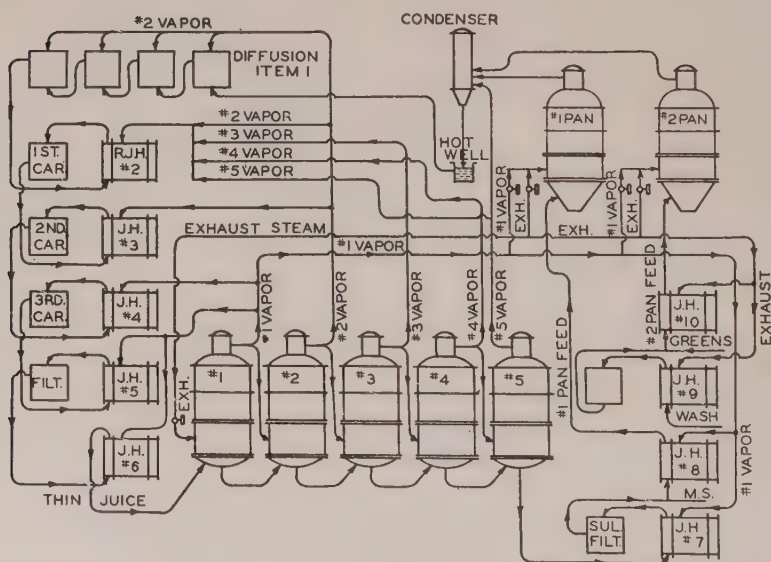


FIG. 11.—Cycle "3-C"—Diagrammatic Sketch Quintuple—Vapor Heaters—Vapor Pans.

Cycle "3-B." Cycle "3-B" is based on operation with the quintuple effect, in which the use of vapors has been made for heating at every possible opportunity within the limits of practice. It is similar to cycle "2-B" discussed previously, and therefore does not need much elaboration here.

There follows a table, diagrammatic sketch (Fig. 10), and detailed heat balance, as in all other cases.

The total steam consumption is listed in the final summary sheet for quintuple effects, and amounts to 65.79 per cent. of the weight of the beets sliced in pounds of steam "from and at 212° F.," as in all other cases. With the customary allowance for contingencies, this becomes 75.79 per cent., which is quite a material reduction as compared with the previous case of the straight quintuple whose consumption was 89.95 per cent.

Cycle "3-C." Cycle "3-C" is the same as cycle "3-B," except that the vacuum pans are operated on vapor instead of live or exhaust steam.

There is attached as in the other case, a complete heat balance for the operation of the quintuple effect, as well as a table of heating and other operations, and the diagrammatic sketch (Fig. 11) showing the disposition of the apparatus, and its functions.

The steam consumption for the factory operating according to this cycle is given in the summary sheet for quintuple effects, at the end of this discussion. The total steam consumed is 61.74 per cent. of the weight of the beets sliced in pounds "from and at 212° F." This amount becomes 71.74 per cent. when the conventional allowance is made for contingencies.

VAPORS AND STEAM FOR PANS AND HEATERS.

VAC. QUINT—VAP. HEATERS—STEAM PANS. CYCLE "3-B."

Item	Station	Weight	B.t.u. Req.	Steam B.t.u. T = 240	No. 1 Vap. B.t.u. T = 230	No. 2 Vap. B.t.u. T = 217	No. 3 Vap. B.t.u. T = 201	No. 4 Vap. B.t.u. T = 179	No. 5 Vap. B.t.u. T = 141
1	Diffusion	B-100,000	5,300,000			5,300,000			
2	Raw juice	140,000	11,330,000			1,110,000 T = 194	3,080,000 T = 186	5,320,000 T = 164	1,820,000 T = 126
3	1st carb.	159,800	3,520,000			3,520,000 T = 194			
4	2nd carb.	152,000	4,100,000		4,100,000 T = 212				
5	3rd sat.	152,000	2,130,000		2,130,000 T = 212				
6	Thin juice	152,000	3,040,000		3,040,000 T = 220				
7	Thick juice	30,600	1,680,000		1,680,000 T = 195				
8	Syr. and M. S.	34,700	1,560,000		1,560,000 T = 195				
9	Wash.	13,600	1,086,000	1,086,000					
10	Greens	16,000	1,280,000	1,280,000					
Total for heaters			35,026,000	2,366,000	12,510,000	9,930,000	3,080,000	5,320,000	1,820,000
12	No. 1 pan	52,100	15,450,000	15,450,000					
13	No. 2 pan	16,000	5,165,000	5,165,000					
Total for pans			20,615,000	20,615,000					

THE HEAT BALANCE IN THE BEET SUGAR FACTORY 331

CYCLE "3-B"—HEAT BALANCE QUINTUPLE EFFECT—VAPOR HEATING.

Body	Specifications	Heat	Juice
(1)	Steam, 42,900 lbs. @ 240 = $42,900 \times 952.1 =$	40,800,000	152,000
	Deduct for heat. $152,000 \times (230.5 - 220 = 10.5) =$	1,597,000	
	Available for evaporation	39,203,000	
	L @ 230 = 958.7. $E_1 = \frac{39,203,000}{958.7} =$		40,900
	Transferred to No. 2		111,100
(2)	Vapor ex. No. 1	39,203,000	
	Add flash, $111,100 \times (230.5 - 217.5 = 13) =$	1,440,000	
	Available heat	40,640,000	
	Deduct for vapor heating (see chart).....	12,510,000	
	Available for evaporation.....	28,133,000	
	L @ 217 = 966.5, $E_2 = \frac{28,133,000}{966.5} =$		29,100
	Transferred to No. 3		82,000
(3)	Vapor ex. No. 2	28,133,000	
	Add flash, $82,000 \times (217.5 - 202 = 15.5) =$	1,270,000	
	Available heat	29,403,000	
	Deduct for vapor heating (see chart).....	9,930,000	
	Available for evaporation.....	19,473,000	
	L @ 201 = 977.2, $E_3 = \frac{19,473,000}{977.2} =$		19,900
	Transferred to No. 4		62,100
(4)	Vapor ex. No. 3	19,473,000	
	Add flash, $62,100 \times (202 - 180.5 = 21.5) =$	1,330,000	
	Available heat	20,803,000	
	Deduct for vapor heating (see chart).....	3,080,000	
	Available for evaporation.....	17,723,000	
	L @ 179 = 990.5, $E_4 = \frac{17,723,000}{990.5} =$		17,900
	Transferred to No. 5		44,200
(5)	Vapor ex. No. 4	17,723,000	
	Add flash, $44,200 \times (180.5 - 148 = 32.5) =$	1,435,000	
	Available heat	19,158,000	
	Deduct for vapor heating (see chart).....	5,320,000	
	Available for evaporation	13,838,000	
	L @ 141 = 1012.6, $E_5 = \frac{13,838,000}{1012.6} =$		13,600
	Thick juice ex. No. 5.....		30,600
Condenser			
	Vapor ex. No. 5	13,838,000	
	Deduct for vapor heating (see chart).....	1,820,000	
	Heat to condenser	12,018,000	
	L @ 141 = 1012.6, vapor to condenser = $\frac{12,018,000}{1012.6} =$..		11,870

VAPORS AND STEAM FOR PANS AND HEATERS.
 VAC. QUINT—VAPOR HEATERS—VAPOR PANS, CYCLE 3-C.

Item	Station	Weight	B.t.u. Req.	Steam B.t.u. T = 240	No. 1 Vap. B.t.u. T = 230	No. 2 Vap. B.t.u. T = 217	No. 3 Vap. B.t.u. T = 201	No. 4 Vap. B.t.u. T = 179	No. 5 Vap. B.t.u. T = 141
1	Diffusion	B-100,000	5,300,000			5,300,000			
2	Raw juice	140,000	11,330,000			1,110,000 T = 194	3,086,000 T = 186	5,320,000 T = 164	1,820,000 T = 126
3	1st carb.	159,800	3,520,000			3,520,000 T = 194			
4	2nd carb.	152,000	4,100,000		4,100,000 T = 212				
5	3rd sat.	152,000	2,130,000		2,130,000 T = 212				
6	Thin juice	152,000	3,040,000		3,040,000 T = 220				
7	Thick juice	30,600	1,680,000		1,680,000 T = 195				
8	Syr. and M. S.	34,700	1,560,000		1,560,000 T = 195				
9	Wash.	13,600	1,086,000	1,086,000					
10	Greens	16,000	1,280,000	1,280,000					
Total for heaters			35,026,000	2,366,000	12,510,000	9,930,000	3,080,000	5,320,000	1,820,000
12	No. 1 pan	52,100	15,450,000		15,450,000				
13	No. 2 pan	16,000	5,165,000		5,165,000				
Total for pans			20,615,000						

THE HEAT BALANCE IN THE BEET SUGAR FACTORY 333

CYCLE "3-C"—HEAT BALANCE, QUINTUPLE, VAPOR HEATERS, VAPOR PANS.

Body	Specifications	Heat	Juice
(1)	Steam, 60,500 lbs. @ 240 = $60,500 \times 952.1 =$	57,600,000	152,000
	Deduct for heat. $152,000 \times (230.5 - 220 = 10.5) =$	1,597,000	
	Available for evaporation.....	56,003,000	
	L @ 230 = 958.7 , $E_1 = \frac{56,003,000}{958.7} =$		58,400
	Transferred to No. 2.....		93,600
(2)	Vapor ex. No. 1.....	56,003,000	
	Add flash, $93,600 \times (230.5 - 217.5 = 13) =$	1,215,000	
	Available heat.....	57,218,000	
	Deduct for pans and heaters (see chart).....	33,125,000	
	Available for evaporation.....	24,093,000	
	L @ 217 = 966.5 , $E_2 = \frac{24,093,000}{966.5} =$		24,900
	Transferred to No. 3.....		68,700
(3)	Vapor ex. No. 2.....	24,093,000	
	Add flash, $68,700 \times (217.5 - 202 = 15.5) =$	1,065,000	
	Available heat.....	25,158,000	
	Deduct for vapor heaters (see chart).....	9,930,000	
	Available for evaporation.....	15,228,000	
	L @ 201 = 977.2 , $E_3 = \frac{15,228,000}{977.2} =$		15,600
	Transferred to No. 4.....		53,100
(4)	Vapor ex. No. 3.....	15,228,000	
	Add flash, $53,100 \times (202 - 180.5 = 21.5) =$	1,140,000	
	Available heat.....	16,368,000	
	Deduct for vapor heating (see chart).....	3,080,000	
	Available for evaporation.....	13,288,000	
	L @ 179 = 990.5 , $E_4 = \frac{13,288,000}{990.5} =$		13,400
	Transferred to No. 5.....		39,700
(5)	Vapor ex. No. 4.....	13,288,000	
	Add flash, $39,700 \times (180.5 - 148 = 32.5) =$	1,290,000	
	Available heat.....	14,578,000	
	Deduct for vapor heating (see chart).....	5,320,000	
	Available for evaporation.....	9,258,000	
	L @ 141 = 1012.6 , $E_5 = \frac{9,258,000}{1012.6} =$		9,100
	Thick juice ex. No. 5.....		30,600
Condenser	Vapor ex. No. 5.....	9,258,000	
	Deduct for vapor heating (see chart).....	1,820,000	
	Heat to condenser.....	7,438,000	
	L @ 141 = 1012.6 , vapor to cond. = $\frac{7,438,000}{1,012.6} =$		7,340

Summary of Quintuple Effect Analyses. The summary sheet and diagram (Fig. 12) accompanying show vividly the results of the consideration of the advisable combinations of quintuple effect apparatus in the beet sugar factory studied in the preceding pages. Further elaboration is hardly necessary, as the picture is plain enough.

The same remarks apply in this case as in the case of the quadruple

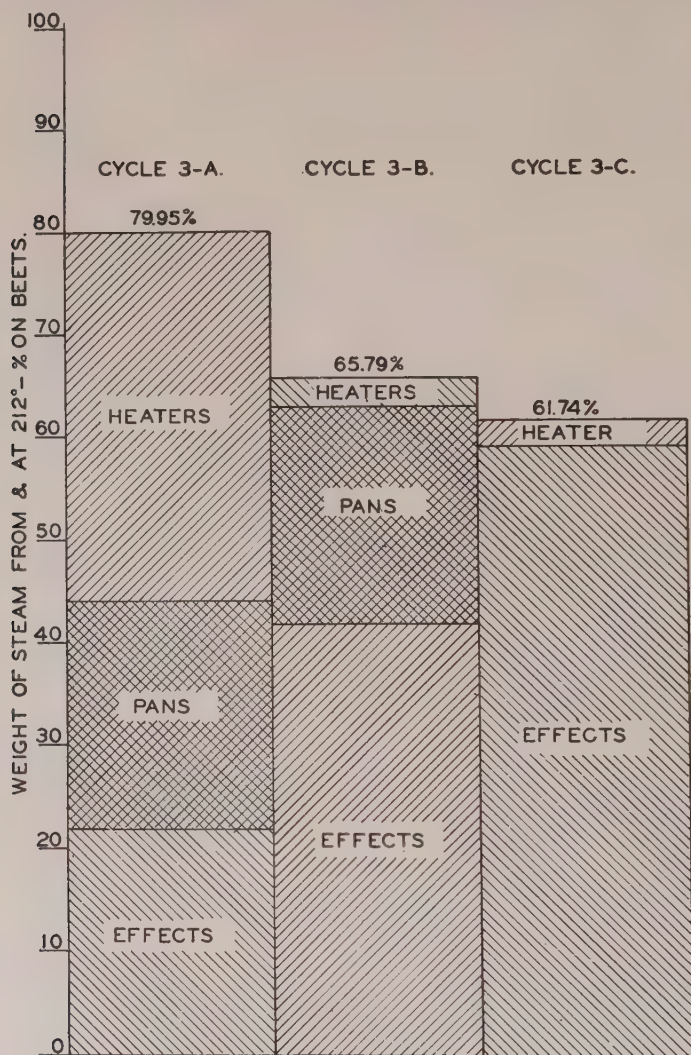


FIG. 12.—Chart Showing Comparative Steam Consumption—Quintuple Effect Combinations.

effect. There are other combinations possible. The local conditions must be taken into consideration in all this work, and these will limit or amplify the possibilities as the case may be. The arrangements suggested under the premises are probably well within the limits of practice providing apparatus is selected which will perform according to prescription.

SUMMARY SHEET.

SCHEDULE OF HEAT CONSUMPTION—QUINTUPLE EFFECTS

CYCLE "3-A"—STRAIGHT QUINTUPLE.

Apparatus	B.t.u.'s Required	Steam, from and at 212° F.	Steam Per Cent. B.
Heaters	35,026,000	36,100	36.10
Pans	20,615,000	21,250	21.25
Evaporators	21,950,000	22,600	22.60
Total	77,591,000	79,950	79.95

CYCLE "3-B"—QUINTUPLE AND VAPOR HEATING.

Apparatus	B.t.u.'s Required	Steam, from and at 212° F.	Steam Per Cent. B.
Heaters	2,366,000	2,435	2.44
Pans	20,615,000	21,250	21.25
Evaporators	40,800,000	42,100	42.10
Total	63,781,000	65,785	65.79

CYCLE "3-C"—QUINTUPLE, VAPOR HEATING, VAPOR PANS.

Apparatus	B.t.u.'s Required	Steam, from and at 212° F.	Steam Per Cent. B.
Heaters	2,366,000	2,435	2.44
Pans			
Evaporators	57,600,000	59,300	59.30
Total	59,966,000	61,735	61.74

Pressure Evaporation.

Preliminary. The fourth group of this series includes the consideration of the so-called "pressure system" of multiple effect evaporation as applied to the beet sugar factory. This method of operation has been put into successful practice in Europe during the past few years.

The usual arrangement consists of either a double effect, or a triple effect operating entirely under pressure, and without condenser. Presumably, all the vapors leaving the various bodies of the apparatus are used either for heating operations, or for vacuum pans, and thus none are wasted. The economy therefore comes in using steam through more stages than has been done before.

This method is made possible through the use of higher exhaust pressures than have been prevalent in the past. It must also be remembered that whereas the total temperature fall is necessarily smaller, on account

of the higher temperatures, not only is the prevailing coefficient of heat transfer greater, but the viscosity of the thick juice is reduced, and the retarding effects of high densities on heat transmission is also reduced, permitting much easier transfer of heat. The exhaust pressures usual for this work are from 17 lbs. to 20 lbs. gauge.

One of the points about which there was considerable doubt at first was whether or not the necessarily high temperatures prevailing in the evaporator would bring about serious losses of sugar by inversion. It is well known that such losses do occur in all heating operations when high temperatures are maintained for appreciably long periods of time, particularly if the juices so heated are in an acid condition. In actual practice, however, by properly designing the evaporator, the time can be so reduced that inversion is negligible.

In 1913, the author tried out this same scheme in the cane sugar industry, but without much success. The obstacle was the fact that the juices to be evaporated in this particular factory had been sulphured, limed, and defecated at 212° F., and upon being heated above this temperature in the evaporator, immediately deposited heavy lime scale on the heating surfaces of the apparatus. In this way, there was a very rapid depreciation of heat transmission, making successful operation difficult and laborious. Without the use of sulphur, the scheme would no doubt have been successful, as shown by similar performances in pre-evaporators.

In order to permit successful operation of pressure evaporators, the power units and boilers must be designed for the work, as the considerable increase of the back pressure will in turn greatly increase the water rate of the turbines and power units, possibly beyond the ability of the steam-consuming apparatus to utilize all of the exhaust, which would vitiate the whole project. High initial steam pressures, and even superheat, are advisable, if not indispensable.

In the cane sugar industry, this radical departure is hardly necessary, for it is easily possible to make a balance by the use of the bagasse fuel alone under the average conditions. In the beet sugar industry, however, where all fuel must come from the outside, any change that materially reduces the steam consumption is justified, *provided it can pay its own way*. It is hoped that the analyses that follow will throw light upon this feature.

Operating Conditions. It is assumed that both the pressure doubles, and the pressure triples will be supplied with exhaust steam at approximately 18 lbs. gauge pressure, corresponding to a temperature of 253° F. It is also assumed that the last body of the evaporator will operate at a slight pressure, so that the temperature of the vapors leaving it will be at 215° F. The distribution of temperatures and pressures, as well as the boiling points, and the value of the corresponding latent heat in B.t.u.'s per pound are given below for both cases.

To delve into the various combinations that are possible with these two sets, would unnecessarily complicate the discussion. There is submitted one analysis for the pressure double, and one for the pressure triple, in which every opportunity has been taken to use vapors instead of steam for heating and vacuum pan work.

THE HEAT BALANCE IN THE BEET SUGAR FACTORY 337

SCHEDULE OF VAPOR AND JUICE TEMPERATURES AND LATENT HEATS FOR PRESSURE EVAPORATORS.

Basis, initial Brix, 12.9; final Brix, 64.0.

System	Nomenclature	Steam Belt 1	Vapor Belt 1	Vapor Belt 2	Vapor Belt 3
Pressure Double Effect.	Vapor temperature ...	253	237	215	
	Boiling temperature ..	...	238	222	
	Latent heat	943.3	954.1	968.4	
Pressure Triple Effect.	Vapor temperature ...	253	242	231	215
	Boiling temperature ..	...	243	233	222
	Latent heat	943.3	950.7	958.1	968.4

HEATING AND EVAPORATING OPERATIONS FOR 100,000 LBS. BEETS PER HOUR. (4) PRESS EVAPORATORS.

Item	Station	Material	Wt. Lbs. per Hr.	Temperature Range			B.t.u. Required
				From F.	To F.	Rise F.	
1	Diffusion	Beets and water	100,000	50	115	65	5,300,000
2	Raw juice	Juice	140,000	113	194	81	11,330,000
3	1st. carb.	Juice	159,800	172	194	22	3,520,000
4	2nd. carb.	Juice	152,000	185	212	27	4,100,000
5	3rd. saturation	Juice	152,000	198	212	14	2,130,000
6	Thin juice	Juice	152,000	200	227	27	4,110,000
7	Ex. evaporators	Thick juice	30,600	This disappears			
8	Syr. and M. S.	Sulph. thick juice	34,700	150	195	45	1,560,000
9	Wash.	Dil. wash.	13,600	115	195	80	1,086,000
10	Greens	High greens	16,000	115	195	80	1,280,000

Item	Station	Material	Wt. Lbs. per Hr.	Bx. Temp.	Bx. In	Bx. Out	Evapo-ration
11	Evaporators	Juice	152,000	227	12.9	64.0	121,400
12	No. 1 vac. pans	Thick juice, etc.	52,100	195	68.0	92.0	13,200
13	No. 2 vac. pans	Greens, etc.	16,000	195	68.0	94.0	4,400

Pressure Double Effect. The quantities and various operations in the use of the pressure double effect are practically the same as in the previous cases, as shown in the tabulated list herewith. Item 6 is again raised to correspond to the increased boiling temperature in the first effect. In view of the fact that the thick juice leaves the evaporator at 222° F. in both the triple and double, item 7, which involves the heating of the thick juice, disappears. Regarding the selection of vapors for these various operations, they hardly need further elaboration, as the same principles set forth previously are followed, therefore, the usual table showing these selections will be submitted.

There is one exception to the above upon which explanations are in order. In the operation of the pans, it has been necessary to split up their requirements between vapors from the first and second effects, whose temperatures are 237° F., and 215° F., respectively. The reason for this

VAPORS AND STEAM FOR PANS AND HEATERS.

PRESSURE DOUBLE EFFECT. CYCLE "4-A."

Item	Station	Weight	B.t.u. Req.	Steam B.t.u. T = 253	No. 1 Vap. B.t.u. T = 237	No. 2 Vap. B.t.u. T = 215
1	Diffusion	B-100,000	5,300,000			5,300,000
2	Raw juice	140,000	11,330,000			11,330,000
3	1st. carb.	159,800	3,520,000			3,520,000
4	2nd. carb.	152,000	4,100,000		4,100,000	
5	3rd. sat.	152,000	2,130,000		2,130,000	
6	Thin juice	152,000	4,110,000		4,110,000	
7	Ex. evap.	30,600	Disappears.			
8	Syr. and M. S.	34,700	1,560,000			1,560,000
9	Wash.	13,600	1,086,000	1,086,000		
10	Greens	16,000	1,280,000	1,280,000		
12	No. 1 pan	52,100	15,450,000		5,000,000	10,450,000
13	No. 2 pan	16,000	5,165,000		1,500,000	3,665,000
Total				2,366,000	16,840,000	45,825,000

CYCLE "4-A." HEAT BALANCE—PRESSURE DOUBLE EFFECT.

Body	Specifications	Heat	Juice
(1)	Steam, 71,900 lbs. @ 253 = $71,900 \times 943.3 =$	67,800,000	152,000
	Deduct for heating, $152,000 \times (238 - 227 = 11) =$...	1,672,000	
	Available for evaporation.....	66,128,000	
	$L @ 237 = 954.1. E_1 = \frac{66,128,000}{954.1} =$		69,200
	Transferred to No. 2.....		82,800
(2)	Vapor ex. No. 1.....	66,128,000	
	Add flash, $82,800 \times (238 - 222 = 16) =$	1,320,000	
	Available heat.....	67,448,000	
	Deduct for pans and heaters (see chart).....	16,840,000	
	Available for evaporation.....	50,608,000	
	$L @ 215 = 968.4. E_2 = \frac{50,608,000}{968.4} =$		52,200
	Thick juice ex. No. 2.....		30,600
(Ex. 2)	Vapors ex. No. 2.....	50,608,000	
	Deduct for heaters and pans (see chart).....	45,825,000	
	Discharged to atmosphere.....	4,783,000	
	$L @ 215 = 968.4. Vapor corr. \frac{4,783,000}{968.4} =$		4,935

Cycle "4-B."

Pressure Triple Effects. The pressure triple effect offers a much better chance to accomplish worth while economies than have heretofore been possible, and is therefore very attractive. The precautions pointed out in the case of the pressure double effect must be carefully observed, as the case is more aggravated.

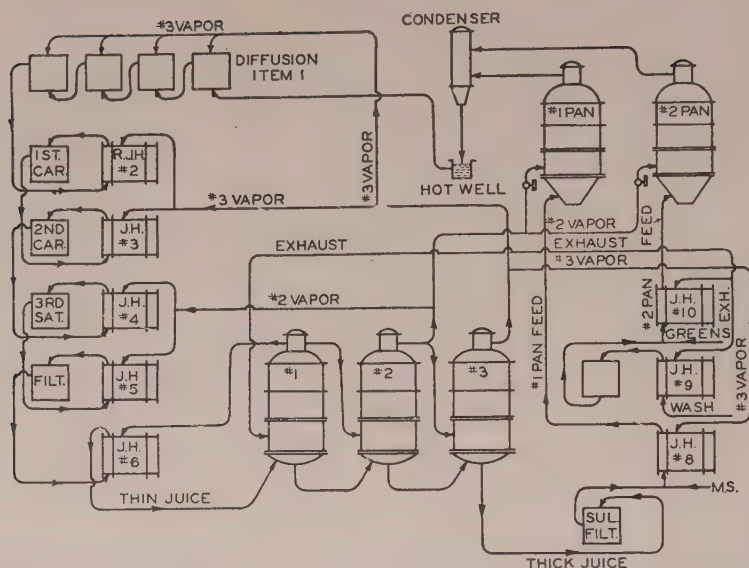


FIG. 14.—Cycle "4-B"—Diagrammatic Sketch Pressure Triple—Vapor Pans—Vapor Heaters.

The diagrammatic sketch (Fig. 14) is given as before, and also a complete heat balance for the work of the evaporator. It will be noticed that in normal operation, there will be a shortage of vapors from the last body

VAPORS AND STEAM FOR PANS AND HEATERS.

PRESSURE TRIPLE EFFECT. CYCLE "4-B."

Item	Station	Weight	B.t.u. Req.	Steam B.t.u. T = 253	No. 1 Vap. B.t.u. T = 242	No. 2 Vap. B.t.u. T = 231	No. 3 Vap. B.t.u. T = 215
1	Diffusion	B-100,000	5,300,000				5,300,000
2	Raw juice	140,000	11,330,000				11,330,000 T = 194
3	1st. carb.	159,800	3,520,000				3,520,000 T = 194
4	2nd. carb.	152,000	4,100,000			4,100,000 T = 212	
5	3rd. sat.	152,000	2,130,000			2,130,000 T = 212	
6	Thin juice	152,000	4,110,000		4,110,000 T = 227		
7	Ex. evap.	This disappears					
8	Syr. and M. S.	34,700	1,560,000				1,560,000
9	Wash.	13,600	1,086,000	1,086,000			
10	Greens	16,000	1,280,000	1,280,000			
12	No. 1 pan	52,100	15,450,000			15,450,000	
13	No. 2 pan	16,000	5,165,000			5,165,000	
Total for pans and heaters				2,366,000	4,110,000	26,845,000	21,710,000

of 1,532,000 B.t.u.'s, which will have to be made up with steam. As a matter of fact, this quantity will be variable as the work of the pan station fluctuates, and it will therefore be necessary to synchronize the work of the factory carefully to conserve heat, otherwise, the shortage will be great at times, and at other times, there will be a surplus which will have to be vented to the atmosphere.

In all these economy schemes, it must be remembered that as one elaborates, one ties up more and more apparatus together, making it increasingly necessary to give careful attention to operation, which under such conditions undoubtedly loses considerably in flexibility.

The steam consumption of the factory is given in the summary sheet, and is 57.64 per cent. on the weight of the beets sliced, in steam "from and at 212° F." With the customary allowance, this becomes 67.64 per cent., which is the lowest of any cycle analyzed, for the hypothetical factory.

Fig. 15 gives a comparison of the performance of the pressure doubles and pressure triples under the conditions assumed previously.

CYCLE "4-B." HEAT BALANCE PRESSURE TRIPLE EFFECT.

Body	Specifications	Heat	Juice
(1)	Steam, 55,100 lbs. @ 253 = 55,100 × 943.3 =	52,000,000	152,000
	Deduct for heating, 152,000 × (243 — 227 = 16) = ...	2,430,000	
	Available for evaporation.....	49,570,000	
	L @ 242 = 950.7. $E_1 = \frac{49,570,000}{950.7} =$		52,150
	Transferred to No. 2.....		99,850
(2)	Vapor ex. No. 1.....	49,570,000	
	Add flash, 99,850 × (243 — 233 = 10) =	998,000	
	Available heat	50,568,000	
	Deduct for vapor heaters (see chart).....	4,110,000	
	Available for evaporation.....	46,458,000	
	L @ 231 = 958.1. $E_2 = \frac{46,458,000}{958.1} =$		48,450
	Transferred to No. 3.....		51,400
(3)	Vapor ex. No. 2.....	46,458,000	
	Add flash, 51,400 × (233 — 222 = 11) =	565,000	
	Available heat	47,023,000	
	Deduct for vapor pans and heaters (see chart).....	26,845,000	
	Available for evaporation.....	20,178,000	
	L @ 215 = 968.4. $E_3 = \frac{20,178,000}{968.4} =$		20,800
	Thick juice ex. No. 3.....		30,600
(Bal)	Vapors ex. No. 3.....	20,178,000	
	Required for vapor heaters (see chart).....	21,710,000	
	Deficit to be made up with steam.....	1,532,000	
	Heat consumption No. 1 above.....	52,000,000	
	Total heat consumption this station.....	53,532,000 B.t.u.'s	

SUMMARY SHEET.

SCHEDULE OF HEAT CONSUMPTION—PRESSURE EVAPORATION.

Cycle "4-A." Pressure Double Effect.

Apparatus	B.t.u.'s Req.	Steam, from and at 212°	Steam, Per Cent. B
Heaters	2,366,000	2,440	2.44
Pans			
Evaporators	67,800,000	69,900	69.90
Total	70,166,000	72,340	72.34

Cycle "4-B." Pressure Triple Effect.

Apparatus	B.t.u.'s Req.	Steam, from and at 212°	Steam, Per Cent. B
Heaters	2,366,000	2,440	2.44
Pans			
Evaporators	53,532,000	55,200	55.20
Total	55,898,000	57,640	57.64

Utilization of Extra Live Steam. In the foregoing discussion, it has been assumed that if the supply of exhaust steam is insufficient for the needs of the factory, operated according to the cycles suggested, this deficiency would be made up by admitting live steam into the exhaust main for the purpose of maintaining the pressure at the desired point.

As has been suggested in the discussion of the cane sugar factories, the amount of heat thus introduced into the system will depend greatly upon the design and conditions of operation of the steam generating plant, and the power producing units, as well as the loads imposed upon the latter. In any event, this quantity of live steam by-passing the power units would be a variable factor subject to much fluctuation in any one factory. It will also be a factor of purely local conditions existing at the particular factory under consideration.

There should be no misunderstanding in the matter of the omission of the Pauly in the preceding discussion, and the reader is therefore requested to refer to the discussion in the chapter on Cane Sugar Factories, cycle "G." He is also referred to the same chapter on the Discussion and Elaboration of the Thermo-Compressor, cycle "M."

Additional economies above those indicated in the consideration of the beet sugar factories can be secured by the use of the Pauly as outlined in cycle "G," or by the use of the thermo-compressor in cycle "M," either of which can be added to any of the beet sugar combinations. As has been fully brought out, the savings possible will be proportional to the amount of live steam used in this way, and can be calculated according to the principles set forth in the above reference.

The preference should be for the thermo-compressor, in that it does not involve the use of excessive temperatures, and can be turned on or off the first body of the multiple effect while in operation, merely involving a slight change in the rate of evaporation at this point, and never a complete shut down, which would be a possibility with the use of the Pauly.

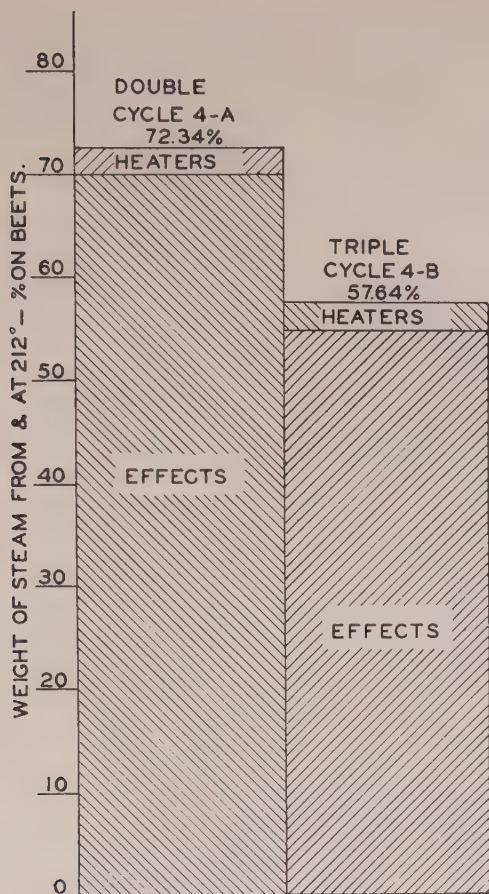


FIG. 15.—Chart Showing Comparative Steam Consumptions—Pressure Evaporators.

The Pauly can be a double effect instead of a single, thus securing greater economy, but involving the danger of still higher temperatures. It is also possible to superimpose the thermo-compressor on the Pauly, getting the economy of both.

If the Pauly is used, the steam will enter its calandria in a superheated condition, and there is considerable advantage in destroying the superheat by the use of a small water spray in the steam pipe between the valve and the evaporator; for it has been confirmed in many instances that the coefficient of heat transmission is very much reduced with superheated steam, for which reason its use should be avoided in evaporating apparatus.

Evaporator Ratings. Below is given what may be considered fair

ratings for the evaporators used in the various cycles analyzed in this chapter operating under the prescribed conditions as outlined.

For juice heaters, the heat transmission can be estimated at from 150 to 200 B.t.u.'s per square foot per hour per degree F. difference between juice and steam, according to steam temperature. For high viscosities, such as thick juice, use only half of these figures.

For multiple effects, the ratings below are given in pounds of actual evaporation per square foot per hour.

Cycle 1, triple effects,	10.0	pounds	per	square	foot	per	hour ;
Cycle 2, quadruple effects,	7.5	"	"	"	"	"	"
Cycle 3, quintuple effects,	6.0	"	"	"	"	"	"
Cycle 4-A, pressure doubles,	6.5	"	"	"	"	"	"
Cycle 4-B, pressure triples,	4.5	"	"	"	"	"	"

These ratings will be subject to variations according to the designs of the evaporators, and their conditions of operation, but they are believed to be safe for good apparatus.

Chapter 4.

Evaporation of Crystallizing Solutions.

The evaporation of a solution from which crystals are continuously formed offers particular problems well worthy of individual consideration. Such problems may be divided into two groups: those in which the viscosity of the mother liquor is so great that the crystals are held in suspension; and those in which they are not held in suspension.

Vacuum Pans. The best example of the first case is the ordinary vacuum pan used in the sugar industry. Here, there is in reality a case of excessively high viscosity, in which the body of the *masse cuite* is so thick, that the settling of the sugar crystals is a practical impossibility. By continuously circulating these crystals, or grains in a supersaturated mass, it is possible to make them grow in size, but not in number, thereby retaining control of a most important factor. This is accomplished by starting with a certain predetermined quantity of small grain, and carefully controlling the feed and boiling temperature, thus practically controlling the growth of these individual crystals without producing new and smaller ones, until the required size is reached. This is the basis of all vacuum pan operation in the sugar industry.

The operation of these vacuum pans has become an art in itself. As the purity of the syrup fed into the pan is lower or higher, successful operation becomes more or less difficult. It is a known fact that crystallization is a slow process, and the speed of growth of these crystals is also proportional to the purity of the solution, being rapid with high purities, and slow with low purities.

All such pan operations are necessarily carried on by batches, more popularly known as strikes. The vacuum pan is merely a single effect evaporator in which the concentration is carried on "dead end."

There are many designs of vacuum pans, some being provided with serpentine coils which usually operate on live steam, as shown in Fig. 1. Others are equipped with steam belts, or calandrias, similar to a standard vertical tube evaporator. Still others have horizontal tubes, both of the latter being designed to work with exhaust steam at from 4 to 15 lbs. gauge pressure. Still other manufacturers have used heating elements of various shapes and designs. Fig. 2 shows a standard calandria pan, which is in practically universal use in most sugar-making countries.

The first essential of a vacuum pan is good circulation. Operation is usually under a vacuum of not less than 23 in., and may run up as high as wanted. It is generally conceded that operation at high vacuum produces "soft grain," and at low vacuum, "hard grain." In refineries, where

hard grain is wanted, low vacuum is used; and in raw sugar factories where hardness is immaterial, and it is a question of exhaustion of mother liquor, high vacuum is used, as the corresponding low temperatures give better exhaustion, besides permitting more rapid operation. It is doubtful that the hardness of grain theory is unassailable, but it is generally accepted in the industry by men of wide experience.

It is very evident that in order to secure uniformity of size of grain, the crystals must be kept in suspension, and thoroughly mixed during the entire strike. This is done by proper arrangement of coils, allowing wide openings at the top and smaller ones at the bottom so that there is an

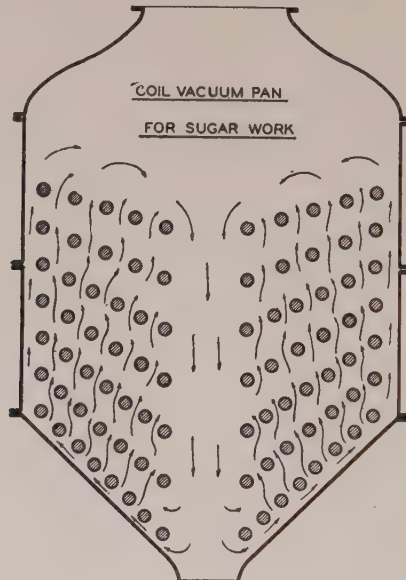


FIG. 1.

uninterrupted passage of the mass, as it becomes swollen with steam bubbles on its way through the heating surface from the bottom to the top. A large free space must be provided in the centre for the return of this mass from the top of the pan to the bottom. (See Fig. 1.)

In calandria pans, large diameter short tubes are used, perhaps 4 in. by 48 in. long. Very large and generous downtakes or return pipes must be provided, sometimes one, sometimes several, according to the size. In some designs, a floating calandria is used, somewhat similar to the "basket" design shown in Fig. 6. There are serious objections to this on account of the difficulty of making repairs, and the excessive strains to which such baskets are subjected by the great weight of the *masse cuite* when the pan is discharged.

In operation, the strike is started by drawing into the pan enough

syrup to cover the heating surface, or at least a good part of it, when evaporation is carried on at a vacuum slightly below standard, until slight supersaturation is reached, when the vacuum is suddenly raised, thus forming a quantity of very small crystals, or "grain." From this point on, the evaporation is carried on at the same vacuum, and the syrup is fed intermittently, so as to supply additional sugar to the strike as fast as the

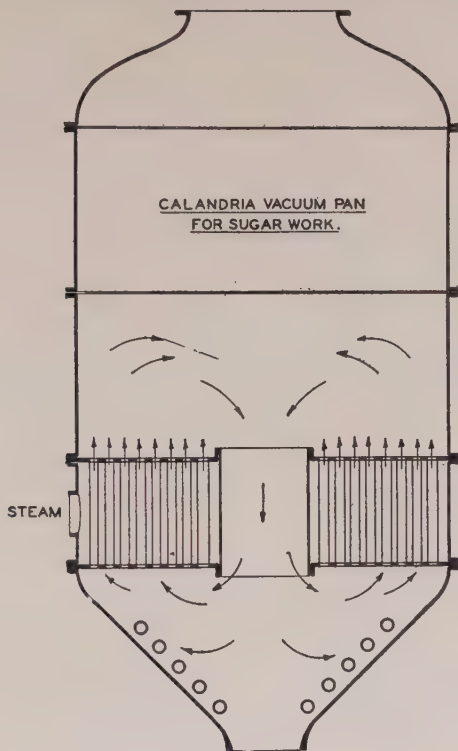


FIG. 2.

crystals can grow. In this way, the level of the liquid in the pan, or *masse cuite*, gradually builds up until the pan is full, which is determined by the level that can be carried without danger of entrainment loss.

As indicated previously, the time required by the strike is dependent upon the speed with which the crystals can grow efficiently. In sugar refineries, very high purities prevail, and the crystal growth is very fast, so that the speed of the pan is governed by the ability of the heating surface to evaporate. Strikes are made in as little as $1\frac{1}{2}$ hours. In raw sugar factories, three crops of crystals or strikes are made: the first, which, on account of high purities, takes a short time, perhaps 2 hours, with good calandria pans; the second, taking from 6 to 8 hours because of reduced

purities, supplemented by a period of post crystallization of say 48 hours in a crystallizer, or horizontal tank wherein the *masse cuite* is kept in a continuous slow motion by means of revolving paddle work; and the third strike, which is very slow because of still lower purities, and which is supplemented by about a week in the crystallizers, the idea being to exhaust the final mother liquor, or "molasses," as far as possible, as this molasses usually brings a relatively low price.

Referring back to the pan designs shown in Figs. 1 and 2, the coil pans usually have about 0.85 to 1.00 sq. ft. of heating surface per cubic foot of strike capacity. For good work, the coils are usually made of 4-in. copper tubing, each individual strand not exceeding 30 to 35 ft. in length. This necessitates a multiplicity of strands to each coil, and it is not uncommon to see coils having two, three, four, six, and even eight strands on very large units.

Calandria pans designed to run on exhaust steam should have about two and one-half times as much surface as coil pans per cubic foot of strike capacity, using 4 to 5 in. tubes about 4 ft. 0 in. long, and down-takes having a net area about one-sixth the total area of the tubes. The capacity of the calandria pans is usually calculated as the net cubic contents when full about 3 to 4 ft. above the top tube sheet, as that is about the height to which the useful strike will settle when ebullition is stopped. It is useless to try to gain much capacity by increasing the height, for after the level reaches a certain point above the top tube sheet, there is a very marked slowing up effect on account of the increased static head.

Low Viscosity Crystallizing Solutions. In most cases, the viscosity of the solutions to be evaporated for salt removal is so low, that it is practically impossible to retain the crystals in suspension and make them grow as has been outlined above. The problem is therefore quite different, and must be attacked from a different angle.

The early pans for this work were patterned along the lines found to be successful in the sugar industry and still are in many cases. The operation, however, cannot be by the batch system described in the preceding pages. Some means must be provided for the removal of the crystals which are continuously formed, and deposited at the bottom. There are, in general, three ways of doing this, as follows:

A. Salt Traps. The bottom of such evaporators are made in the shape of a steep cone, to permit the salts to all settle to the lowest point, having a fairly large outlet. To this outlet are connected two or more "salt traps," or recipients into which the crystals settle, one trap being connected to the evaporator at the time. The size of the traps is determined by the volume of the salts expected in a given time. (See Fig. 3.) There are provided sight glasses through which observations can be made to determine when the traps are full. At the bottom is a grid with backing wire and fine meshed screen, through which the particles of salt cannot pass. Immediately above the screen, located in the side of the trap, is a large hinged door, through which the salt is removed.

When the trap is full, it is cut off from the evaporator, and the other trap turned on. Either air pressure, steam pressure, or both are applied

above the salt level, and the mother liquor, or brine, is forced back through the screen out through the bottom of the trap, through a drain pipe back into the evaporator, as indicated in Fig. 3. Salt can be further dried by forcing air through from the top, and discharging to the atmosphere below. When the pressure is stopped, the door is opened, and the salt removed by hand, with shovels, or by other convenient means.

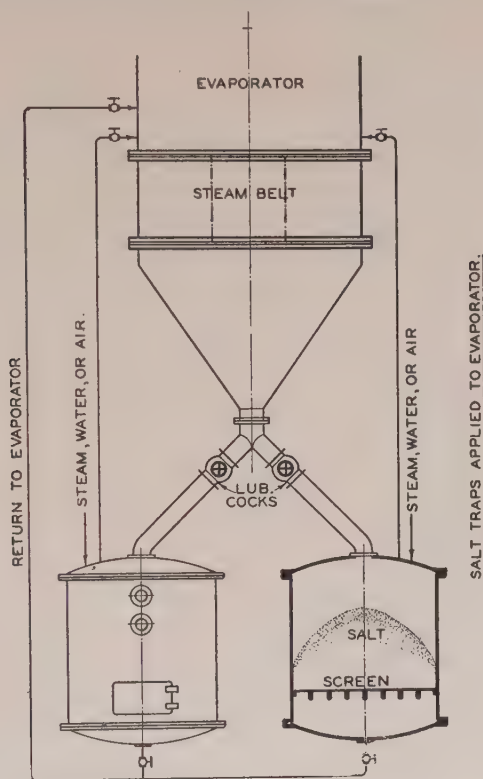


FIG. 3.

Sometimes, the entire bottom of the trap is arranged with a pair of hinges, and made tight with swing bolts, the screen and grid being attached to the bottom. This permits the whole bottom to swing open, allowing the salt to fall into receiving bin or hopper.

It is preferable to make these traps of large diameter and shallow depth, as in this form they drain more quickly without applying excessive pressure, which tends to pack the salt and make its subsequent removal laborious. It is advisable to provide small high pressure water connections to all points likely to become clogged with salt. Thus it is easily possible to dissolve out any troublesome accumulation.

Particular attention must be paid to the various valves through which the mother liquor laden with salts must pass, as they are usually a source of trouble. The most successful device for this is a lubricated cock.

It is advisable to have a fair sized connection from the top of the trap to the evaporator above the upper tube sheet which permits a return flow, particularly where mechanical circulation (discussed later) is used. This creates a circulation from the lower part of the evaporator through the

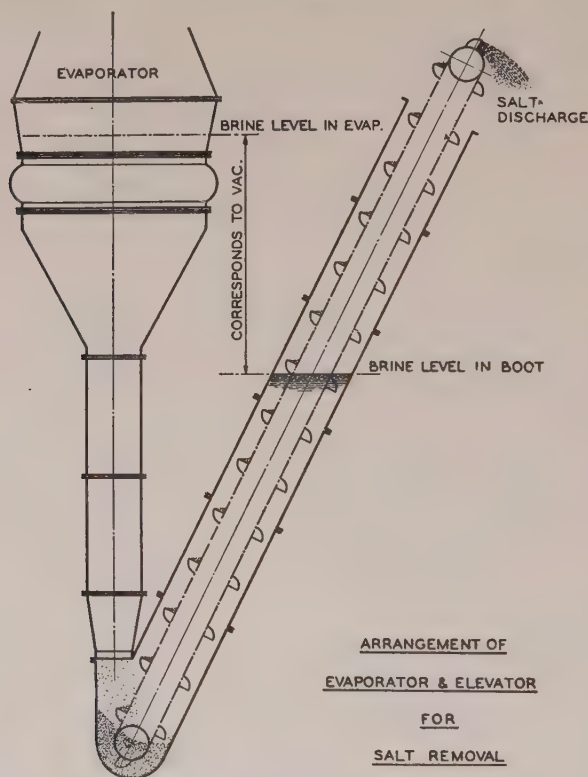


FIG. 4.

trap, where in the prevailing quiescence, the salts drop out, the mother liquor free from salts returning to the evaporator above the top tube sheet. There is a double advantage in this as the contents of the traps, being continually changed, remain at a more uniform temperature.

B. Elevators. A method at one time in extensive use in the salt industry, where very large units are operated, was to use a sealed elevator for the removal of salt from evaporators. This is shown in Fig. 4. The bodies of the apparatus are located at a fairly high elevation, and at the bottom of each is a boot with an elevator having screened buckets, the

whole being enclosed in a watertight casing extending on an angle to one side, clear of the evaporator. The casing extends to such a height that when vacuum is broken, the brine will not overflow at the top. This arrangement is virtually an inverted syphon, the difference of level of the brine in the evaporator as compared with its level in the boot casing forming an equivalent barometric seal against the vacuum existing in the evaporator. The elevator itself extends beyond the casing, and discharges salt from above, the brine being drained out by gravity through the wire screened buckets.

SALT REMOVAL BY PUMPING.

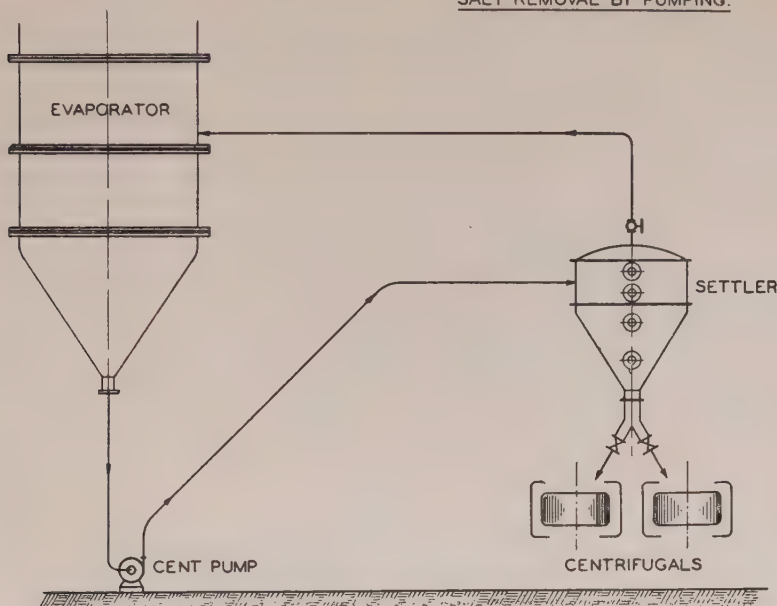


FIG. 5.

For handling sodium chloride, whose solubility varies little with temperature changes, this arrangement has worked with fair success, but for other solutions of materials having a wide range of solubility as the temperature changes, the screens quickly become coated with salt which deposits upon the wires, after which no drainage is possible, the result being that both mother liquor and salt are bailed out of the evaporator, thus making this type of equipment practically useless. In addition, it is expensive in first cost, and requires constant upkeep.

C. Outside Settling. A much more satisfactory arrangement (Fig. 5) is to place a centrifugal pump just below the evaporator, drawing brine and accompanying salt, discharging it into an elevated settling tank, and returning the brine free from salt back to the evaporator. This tank can always be kept by valve manipulation under pressure irrespective of the

vacuum conditions in the apparatus, so that as salt accumulates in the bottom, it is easily seen by observation through sight glasses provided for this purpose, it can be withdrawn either directly into the centrifugals for drying, or into a small mixer feeding the centrifugals. This makes a very satisfactory arrangement, and keeps the evaporator fairly clear of sus-

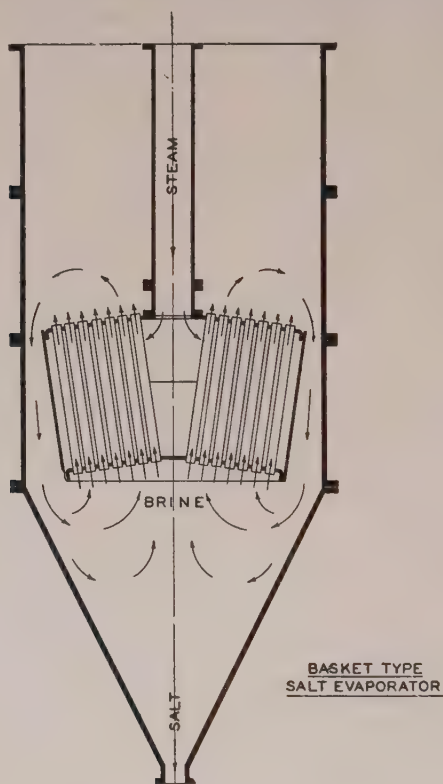


FIG. 6.

pended salts at all times. It is manifestly applicable only to large operations. For very small units, the salt trap is preferable.

The brine from the centrifugals is returned to the evaporator either directly or through the regular feed. It has been shown that the material coming out of the bottom of the settler mentioned above will carry from 30 to 40 per cent. salts, the balance being brine. No doubt, there are certain unavoidable heat losses, for which allowance must be made, but with proper insulation, such losses can be kept down to very reasonable proportions, certainly less than they would be in an elevator installation, whose work must be supplemented with centrifugals in any case.

Mechanical Circulation. It has been found in practically all cases of evaporation of low viscosity salting solutions, that there is an ever present tendency to deposit some of the salts upon the heating surface of the evaporator. In such operations, the method employed to overcome this tendency has been to carry high levels, and use fast liquor circulation. The evaporator must be provided with large downtake area and short tubes of rather large diameter. The type in most common use for this work is the so-called basket type shown in Fig. 6, or some modification of it. Here, there is a closed steam drum with steam inlet in the centre, and an annular downtake, between the drum and the cylindrical body of the evaporator, permitting free passage of the mother liquor from the top to the bottom of the evaporator. This apparatus gives fairly good service, but in spite of all efforts, there is an inevitable salting of surfaces.

In order to overcome this, a radically different method is used, whereby the liquor circulation is formed mechanically by the use of a propeller, or some sort of pump in the central downtake. This method has been fully discussed in the chapter on Mechanical Circulation, to which the reader is referred.

As has been pointed out previously, some of the problems in evaporation become very complicated, especially when the solution contains various salts which must not only be removed from solution, but separated as well, and the field is so broad as to be beyond the scope of this book. It will always be necessary to keep in mind basic principles for the successful solutions of such problems, and it is hoped that these are made clear in this text.

Chapter 5.

Extracts and Dyes.

Under the heading of extracts and dyes, is included a general discussion of the problems encountered in the concentration of tan bark extract, chestnut extract, logwood extract, and similar materials.

All these solutions are derived from the leaching of finely divided chipped wood or bark, this operation being done either in open wooden tanks, or closed copper autoclaves working under pressure. If open tanks are used, the specific gravity of the solution will be from 10 to 15 Barkometer; if autoclaves, it will be about twice this density, therefore requiring only half the amount of evaporation to reach the final product, the specific gravity of which will be between 45 and 55 Twaddell. It will be seen, therefore, that the ratio of evaporation in any case is very high. Assuming the probable average case, in which the range of densities will be from 12 Barkometer to 48 Twaddell, the specific gravities are 1.012 and 1.24, respectively (see Sec. I, Chapter 1), and the ratio of the initial to the final volume will be 20 to 1.

The rating of extract plants is usually in barrels of finished extract per day, each barrel containing about 50 gall. Thus, a 100-bbl. plant would mean one which produces 5000 gall. per day of finished extract, having been concentrated from an original of 100,000 gall. per day, or 4167 gall. per hour, corresponding to 35,200 lbs. per hour. The evaporation from 20 to 1 concentration is 3960 gall., or 35,000 lbs. per hour.

In handling all such materials, three precautions are necessary, as follows: (1) the solution must not come into contact with iron or steel at any point, which causes serious discolorization of the product; (2) it must not be overheated, as high temperatures seriously injure it. Where heating is unavoidable, the time element must be maintained as brief as possible. (3) Care must be used to avoid the introduction of lime and magnesium salts into the leach.

It is necessary, in the leaching process to use pure water free from iron and scale. Iron present in this water even in small quantities will ruin the product. Lime salts will combine with the extract, forming an insoluble compound with corresponding loss.

It is therefore customary to use either copper or aluminum construction or wooden tanks throughout, and it would be most advisable to use surface condensers with the evaporator in order to save all the water removed from the solution, thus making it available for re-use as leach. Furthermore, this acts as a double check against entrainment losses due to care-

lessness on the part of the operators. The condensed vapors alone will not suffice for the complete leaching of the chips, for a considerable amount of water remains in the exhausted refuse. The makeup should be with softened, or better, distilled, water.

The extract contains a measurable amount of organic acids from the wood, and the condensate from the evaporator is decidedly acid and very corrosive, even to copper. All parts must therefore be made fairly heavy, otherwise they will be very short lived.

The operation of the evaporator for this work does not involve any particular difficulty. It is not subject to serious foaming, and standard construction can be used. The solution evaporates very easily and rapidly, permitting high ratings, eight pounds per square foot per hour being quite possible in quadruple effects fairly well designed.

Usually, the apparatus is operated on exhaust steam from the various power units in the plant, much the same as in the case of the sugar factory. The chipping of wood requires considerable power, and adding to this the various pumps, conveyors, etc., there is quite a quantity of exhaust steam available, which may or may not be sufficient for the operation of the evaporators, according to local conditions.

Evaporation is usually in quadruple effect, beginning with about 2 to 5 lbs. pressure, and finishing with as high a vacuum as possible, usually 26 in. or better in the winter.

At final concentration, the viscosity is quite high and for good work, it is recommended that the final step from about 30 Twaddell to 48 Twaddell be done in a single effect finisher, which can operate on the continuous batch system, as explained in the chapter on Feeding Systems. It must be remembered that the product is delicate, and very much affected by high temperatures, otherwise more economical concentration could be accomplished by countercurrent operation, or by the use of the continuous batch system on the quadruple effect. This latter is actually done in some plants. In others, the apparatus is run dead end until the concentration reaches the proper density in the last body, when steam is turned off and the contents dumped, and a new batch started again. This latter method is not recommended, as it involves too many interruptions, and an ever varying rate of evaporation as the concentration changes.

It would seem, therefore, that much is to be gained in perfect control and quality, and little to be lost in economy by use of the finishing pan as suggested.

As to the cleaning of the heating surfaces, one good boil-out once a week with about a 10 per cent. solution of soda ash for a period of 10 to 12 hours will give very good results, and will prove satisfactory if care has been used not to introduce scale into the system by the use of raw water in the leach.

For calculating the heat balances, the general method suggested in the chapter on this subject should be followed.

Chapter 6.

Organic and Foamy Materials.

The concentration of organic foamy materials requires particular care and treatment. Usually, it is impracticable to use a standard evaporator. A typical example of this is distillery slop.

Experience through many years of practice has shown that in a non-crystallizing solution, the best method of overcoming the tendency to foam is by the use of long tube evaporators, having relatively high vapor velocities. When the proportions are selected so that speeds not less than 15 feet per second are maintained at the tube outlets, foam disappears altogether, being broken up by the intense beating action caused by the rapid vapor streams. Unfortunately, as has been brought out in the chapter on natural circulation, high vapor speeds are always accompanied by correspondingly low liquor speeds at the lower part of the tubes before evaporation begins, and if there is any tendency to foul, this will be very much aggravated, and will require careful attention.

Control of feed in long tube evaporators is a very difficult matter. It is impossible to accomplish this with any degree of satisfaction by regulating the level of the liquor, for the necessary level to obtain proper circulation is a variable factor, depending upon a number of contributing causes. It will increase or decrease if the apparatus is operated slower or faster, in other words, the faster the rate of evaporation, the lower the required level, and conversely. It will also vary with the density of the vapor in the tubes, the lower the density, the lower the level. This level will also vary as the temperature of the liquor entering the evaporator is reduced to a point below ebullition at the existing pressure conditions. It will also vary with the cleanliness of the tubes, being lower for clean tubes, and higher for dirty tubes.

This control problem has been the cause of failure or rejection of many an apparatus. The eventual solution is not regulation of the level, but of the amount of liquor actually circulating from the bottom to the top of each body of the evaporator. This feature is the subject of existing patents, and is a rather recent development. Suffice to say, however, that the control is so arranged that a fixed quantity in gallons per minute is actually circulated, determined by experimentation as proper to give the desired results. Incidentally, it has also been found that it is unwise to cut down this amount beyond a certain point, for under such conditions the fouling at the bottom of the tubes increases rapidly, and thus slows down the evaporator, although momentarily its initial capacity might have been greater. It has also been well established that the entering feed

should be preferably at or above the boiling temperature in the first body of the multiple effect. If not, this feed should be sprayed into the vapor space for preheating by direct contact with vapor, rather than by contact with the heating surface for which it is not designed. To illustrate, in a certain evaporator, with a given amount of circulation, having tubes 16 ft. long, feed at boiling temperature gave a level of 18 in. above the bottom tube sheet, and at 50° F. below the boiling point, the level rose to 12 ft. above the bottom tube sheet, thus immobilizing a great part of the heating surface.

Another difficult problem encountered in the long tube evaporator is the prevention of entrainment losses. There are, however, recent designs of catchalls capable of accomplishing the task, so that this feature need not be the determining factor.

It very often happens that the vapors given off by these organic solutions contain corrosive gases, and that even with proper venting, which it is unnecessary to emphasize here, the condensate is also of a corrosive nature, particularly in the presence of air. Research work to determine these characteristics is a very wise precaution, and enables one to provide against otherwise unseen contingencies.

Sometimes it is desired to concentrate to fairly high densities, in which case it may be desirable to run the evaporator "counter current," the highest density being in the first body instead of the last. (See chapter on Paper Mill Waste Liquors.) This permits much higher concentration as the viscosity, which is a function of the temperature, is very much reduced, and permits further concentration without more effort. With counter current operation, it is necessary to pump the feed from each body to the preceding one, as the natural difference of pressure is opposite to the desired direction of flow. The discharge of these intermediate pumps should preferably terminate in the vapor space in a spray for direct heating.

There is another compromise method used for high concentration, wherein the solution is fed first into the second effect, and then passed by difference of pressure on to the last body, and pumped to the first effect for final concentration at maximum temperature.

For the relative economy of these schemes, the reader is referred to the chapter on the Heat Balance, in which the general method of making such calculations is fully detailed.

When the solution being evaporated fouls the apparatus very quickly, it is advisable to clean oftener, as it is easier to make two light cleaning operations than one heavy one, where there is danger of the incrustations getting beyond control.

In the operation of all long tube evaporators, it is essential that the pressure and vacuum conditions be maintained as uniform as possible, as such apparatus suffer more from fluctuations than the standard short tube designs.

As to materials of construction the chemical composition of the solution must be taken into consideration for proper selection. For many materials containing acids, copper is well suited. Monel metal, aluminum

and various alloys are also useful. Cast iron stands well, provided there is no direct impingement upon the metal, in which case the products of chemical reaction are eroded, exposing ever new surfaces with rapid disintegration.

If the ratio of evaporation is great, time and expense will be saved by splitting up the concentration of the solution into two stages, the larger range of low densities to be done in multiple effect, and the high range representing a fractional quantity to be done in a single effect finishing pan, especially designed for the work in question.

It will usually be found that in the concentration of organic solutions, the particular body of the multiple effect to suffer most from fouling will probably be next to the last body, rather than the last. Experience has shown this to be the case many times, and contrary to what might be expected, heavy, viscous solutions as a rule do not foul as much as at some intermediate stage of concentration.

If it is attempted to concentrate a foamy solution in an evaporator having sluggish vapor circulation, very serious difficulties will be encountered. The entire vapor space will become filled with foam, and there will be considerable entrainment. The old method was to carry low levels, in order to keep the foam down. This resulted in the breaking of the bubbles against the upper part of the heating surface, which became rapidly coated with a thick layer of the material being concentrated, so tightly baked on as to be almost impossible of removal. Sometimes sprays are used to break the foam, but it is only after one has battled for days with a problem of this sort that one realizes the seriousness of it.

When practice has not been established thoroughly, a preliminary study of the problem should be made, and then the design confirmed or altered by investigation with a pilot plant. As has been mentioned before, care and judgment must be used in arriving at conclusions, for the relations existing between a small and a large apparatus are problematical. Nevertheless, certain determinations can be made with the pilot plant which are discernible in no other way.

Chapter 7.

Paper Mill Waste Liquors.

In all pulp mills, there is a large quantity of waste liquors to be disposed of in some way. It may be, as in the case of the sulphite mills, that these liquors when drained into nearby streams, cause serious contamination and, in many cases, legislation has been passed prohibiting the dumping of such solutions. It therefore becomes necessary to evaporate them to a high concentration, and as such, to either sell them for some useful commercial purpose, or burn them. In other processes, such as the sulphate (kraft) process and the soda process, the waste liquors have to be evaporated and incinerated to recover the valuable reagents they contain, otherwise the process is absolutely impractical from a commercial point of view.

The problems encountered in the evaporation of all paper mill waste liquors are closely related, consisting of the evaporation of a foamy solution, containing both salts and organic matter, this amounting to about 6 to 10 per cent. of the weight of the solution. The desired ultimate concentration required is from 45 to 55 per cent. solids by weight, at which point the viscosity assumes serious proportions. In some instances there is a precipitation of solids during the last stages of this work, although this may not always be the case. Very often, non-condensable gases are given off during evaporation, such gases being possibly of a corrosive nature.

The group is therefore of considerable importance, having many extensive applications, and for this reason is treated in detail here. Usually, there is little or no waste heat available for this work, and so economy is a matter of prime importance. In the case of sulphite liquor, it is quite probable that if the concentrated liquor is burnt in proper furnaces under boilers, sufficient heat will be generated to supply the requirements of the evaporating station, and perhaps some to spare under favorable conditions.

The matter is therefore one requiring careful study with a view of securing maximum steam economy, and it is felt that considerable light will be thrown upon the subject by analyzing a specific problem, which is arbitrarily assumed as follows:

Solution to be treated.....	100,000 lbs. per hour;
Initial density	8 per cent. solids by weight;
Final density	50 per cent. solids by weight;
Evaporation	84,000 lbs. per hour;
Initial temperature of liq.....	175 degrees F.
Steam pressure available.....	10 lbs. gauge;
Vacuum obtainable	27 in. (referred to 30-in. bar.).

The method of attack is the same as has been outlined in other problems discussed in this text. The steam temperature corresponding to the available pressure of 10 lbs. gauge is 240°F. , and that corresponding to the obtainable vacuum of 27 in. is 115°F. , so that the total overall temperature drop from the steam chest of the first body of the evaporator to the vapor side of the last is the difference between these two, or 125°F. , which, as has been pointed out before in the text, is not all available as net temperature difference across the heating surface.

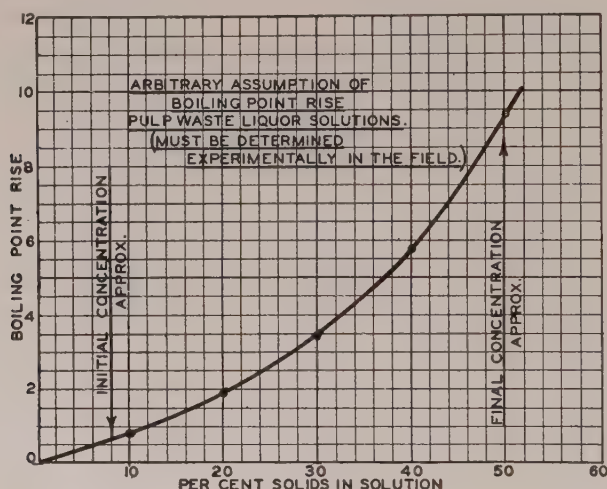


FIG. 1.

There will be a loss of this temperature in each body of the evaporator due to the fact that in every case the temperature of the liquid being evaporated will be greater than that which corresponds to the pressure of the vapor leaving it, this being the familiar boiling point rise discussed many times before. The amount of this loss must be determined experimentally for the solution to be concentrated at the various densities at which it is likely to occur in the various sections of the evaporator. It is assumed in this case that it will vary as shown in Fig. 1, which has been made up arbitrarily, but which probably does not vary greatly from the truth for the type of solution under consideration.

Another loss of temperature drop will result from the inevitable static head, or hydrostatic pressure of the liquid upon the boiling surfaces of the evaporator increasing the boiling point correspondingly, inasmuch as this pressure is superposed upon the vapor pressure existing in the vapor space above the liquid being evaporated. In some apparatus, such as the Lillie, and the Inverted Kestner, the static head has been practically eliminated by the use of pumps which circulate the liquor over the heating surface in thin sheets or films, from which evaporation takes place at substantially vapor pressure. The Moore evaporator is also of this type,

although it is a question whether or not the static head has been completely eliminated, particularly in the last body of the series. In all such cases, the cost of producing this circulation as well as the complications involved must be debited. It must also be remembered that it is almost impossible to circulate equal amounts of liquor over the entire heating surface, and that failure to do so must inevitably result in excessive local concentration in some parts and correspondingly inadequate evaporation in others, these two making up a general average representing the work of the evaporator. Such operation would not seem conducive to satisfactory

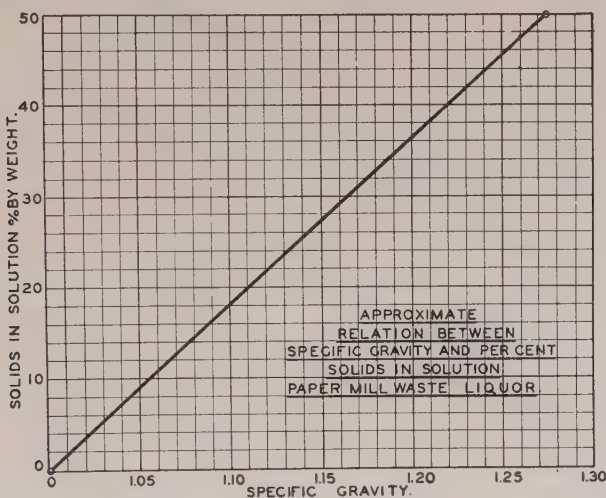


FIG. 2.

performance, particularly in the body in which ultimate concentration is carried on.

For the purpose of this analysis, it will be assumed that the average static head upon the heating surfaces is 12 in. of liquor, and allowance will be made accordingly. Fig. 2 shows the relation assumed to exist between the specific gravity of the solution and the percentage of solids in solution, which we believe closely approximates the truth for the case in hand.

Fig. 3 is a chart by means of which it is possible quickly to determine the amount of temperature drop lost by static head, when the head is known, as well as the specific gravity of the solution, and the temperature of the vapor at which evaporation is taking place.

Both the B.P.R. loss as shown in Fig. 1 and the static head loss determined from Fig. 3 must be subtracted from the total temperature drop to give the net working drop upon the heating surfaces of the contemplated evaporator, and this for the conditions existing in each individual body.

As regards the depreciating effect of viscosity of solution, as has been pointed out in other parts of this book, this is a very elusive factor, and

one upon which there remains considerable investigation in process. It is positively known, however, from the work that has been done, that it is a most serious factor. Figs. 4 and 5 show the relation between viscosity and heat transmission, calculated from data contained in Hugh K. Moore's papers on the evaporation of sulphite liquor, and this information will be

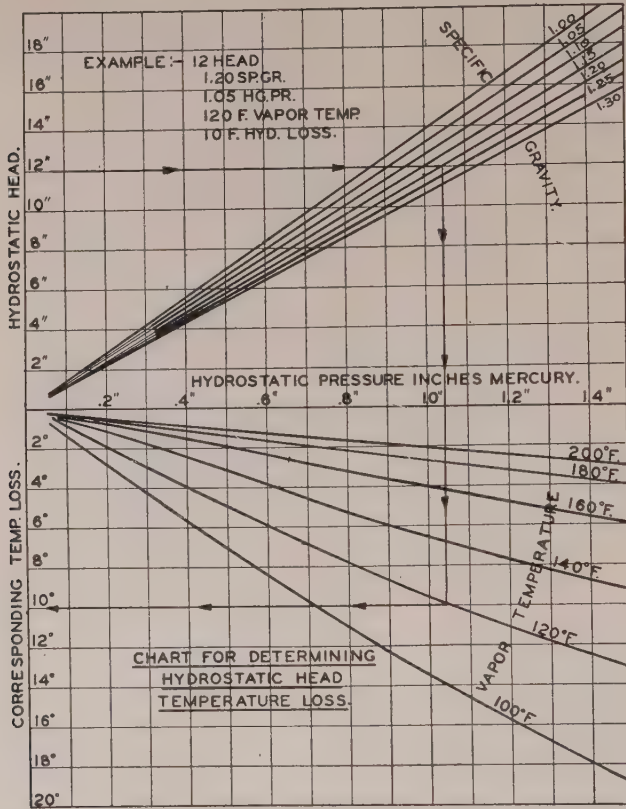


FIG. 3.

used below in determining the conditions of operation of the apparatus under consideration.

Moore also gives a curve showing the viscosity of sulphite liquor for various densities and temperatures, which is reproduced here as Fig. 6. As is well known, it is brought out in this curve that the viscosity decreases greatly as the temperature increases, and it is evident from this consideration that there is much to be gained in capacity and ease of operation by carrying high concentration at high temperatures. There can be no objection to this, inasmuch as the solutions in question are not adversely affected thereby. It would seem, therefore, that there are,

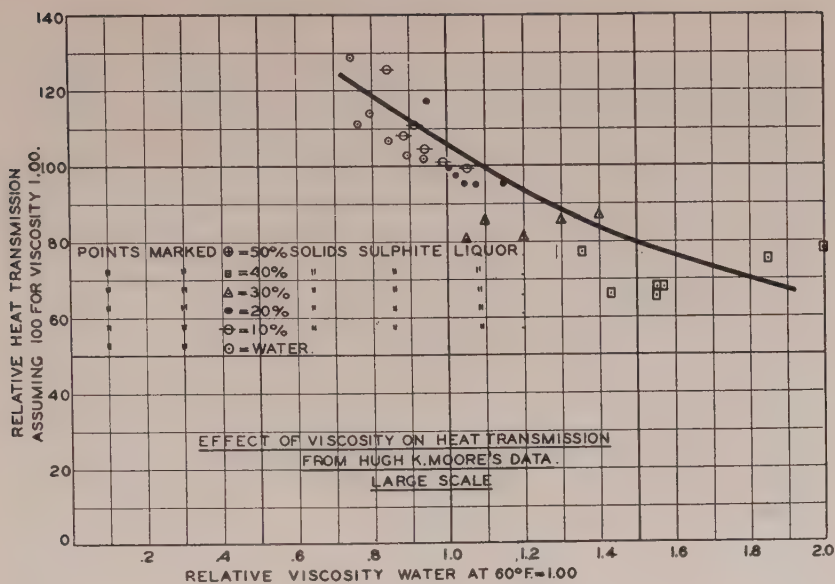


FIG. 4.

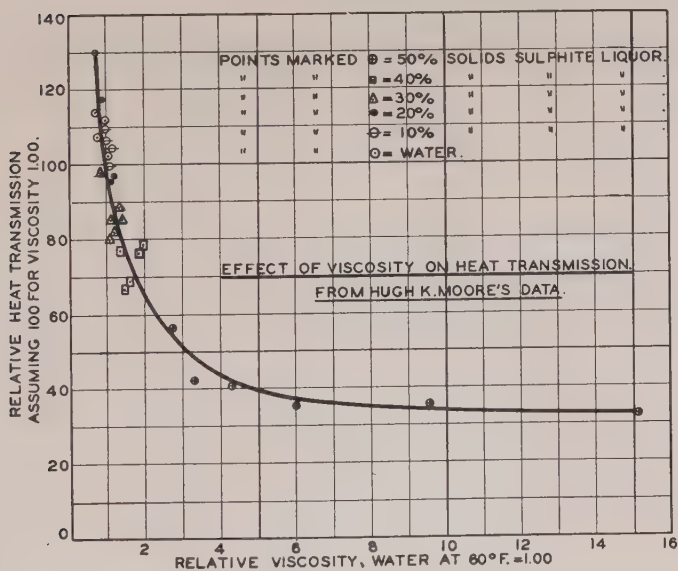


FIG. 5.

among others, three ways to successfully operate a multiple effect under the premises.

1. It can be operated in parallel current (the standard method) in which feed enters the first body and goes forward to the second, third,—last, at as high a concentration as practicable, without encroaching excessively upon the capacity of the apparatus, the ultimate and final concentration being carried on single effect in a finishing pan, whose vapor

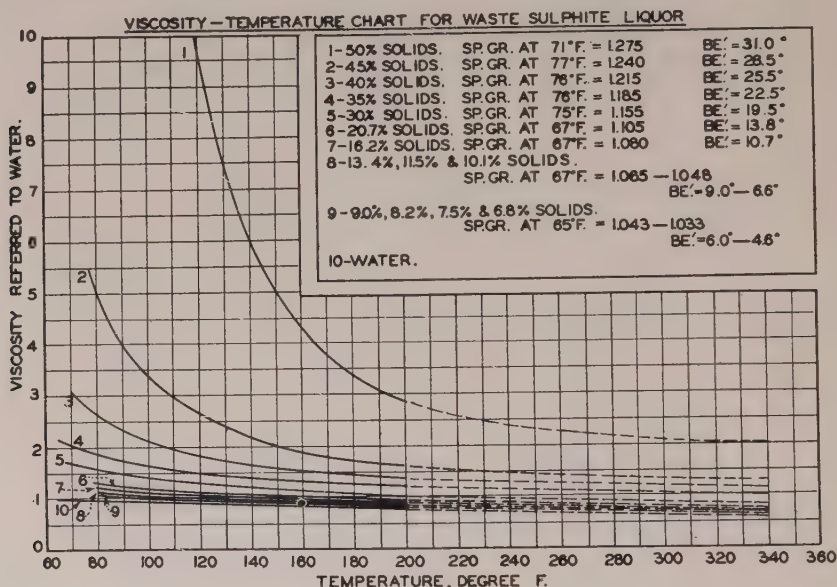


FIG. 6.

space is maintained at a sufficiently high temperature to minimize the ill effects of viscosity. This is probably the simplest method.

2. It can also be operated counter current, where the feed enters the last body, and is pumped back from effect to effect, ultimate concentration taking place in the first body of the series at maximum temperature.

3. The third method is to feed the liquor into the second body first, and go forward from there up to the last effect, from which the partially concentrated liquor is pumped back to the first body for final and ultimate concentration at maximum temperature.

4. It is not advisable to run completely parallel current without the finishing pan, as the viscosity in the last body at relatively low temperatures will be very great, and will require large surfaces in proportion to the work, and besides will give lots of trouble. This combination will therefore not be considered in the discussion that follows:

The temperature of the solution arriving at the evaporator has been assumed at 175° F., and in so doing, it is felt that it is not asking too much

to maintain this temperature by proper provisions against radiation, as this factor has quite an influence upon the steam consumption according to any cycle. It pays to spend a little money on heat conservation, as has been brought out many times before. (See chapter on the Heat Balance.) It would introduce too voluminous calculations into this discussion to provide for this temperature as a variable. If the reader is vitally interested, such calculations can be made by following out the scheme shown in the heat balances below.

The proper establishment of the temperature drops for the various cycles suggested is the next step. As to the number of bodies per set in series for the multiple effect, this is a matter for practice to determine. It will probably not pay to increase the multiplicity beyond the point where there is available less than a gross average of about 20° F. for each body in overall temperature difference, as established by the available steam pressure at one end, and the obtainable vacuum at the other. Otherwise, the rate of evaporation per square foot of surface per unit of time will be very low, requiring excessive costly surfaces for the desired capacity. It is therefore the conclusion that a sextuple effect is about the limit commercially for this work. Analyses will be made for the three cycles suggested above, using quadruples, quintuples, and sextuples.

Below is given a table showing a rough approximate of the amount of evaporation in each body of these sets based upon the assumption that for parallel current (1), the evaporation in bodies 1, 2, 3, will be in the proportion of 1.0, 1.1, 1.2. This will also hold good for cycle (3) where the feed enters the second body first. For cycle (2), the counter current operation, it is assumed that the increase will be in the same proportion, but in the reverse order, beginning with the last effect, and ending with the first. It is also assumed that for cycle (1), the finishing pan will evaporate 5000 lbs. per hour, leaving 79,000 lbs. per hour to be evaporated in the multiple effect. These assumptions are made for rough preliminary purposes to permit the determination of the solids in solution in each body, and from this the boiling point rise, and the specific gravity of the partially concentrated solution, all of which data are incorporated in the log.

It is now very evident that until an approximate distribution of temperatures can be made, it will be impossible to determine the correct static head losses for the various bodies, as these are functions of the vapor saturation temperature. Nor will it be possible to determine the probable coefficients of heat transmission, and the viscosity effects, all being dependent upon temperatures. It is therefore necessary to make an intelligent guess as to the temperature distribution, to be rectified afterward when other variables have been properly estimated.

For parallel current operation, it is known that for many solutions, an equal division of pressures often approximates the truth, and the schedule below is based upon this premise. In this log is also shown the probable static head loss, as determined from the specific gravity of the solution (fixed by the percentage of solids in solution and Fig. 2), and the vapor temperatures, assuming 12 inch head. (See Fig. 3 discussed previously.)

TENTATIVE EVAPORATION ESTIMATES.

Body	Spec.	Cycle 1. Parallel Cur.			Cycle 2. Counter Cur.			Cycle 3. Special.		
		Quad.	Quint.	Sext.	Quad.	Quint.	Sext.	Quad.	Quint.	Sext.
(1)	Evap.	17,200	13,200	10,550	23,750	19,600	16,800	18,250	14,000	11,200
	Per cent. solids..	9.5	9.5	9.0	50.0	50.0	50.0	50.0	50.0	50.0
	B. P. R.....	0.8	0.8	0.8	9.5	9.5	9.5	9.5	9.5	9.5
	Sp. gr.	1.05	1.05	1.05	1.28	1.28	1.28	1.28	1.28	1.28
(2)	Evap.	18,900	14,500	11,600	21,900	18,200	15,700	20,100	15,400	12,300
	Per cent. solids..	12.5	11.0	10.0	20.0	22.5	24.0	10.0	9.5	9.0
	B. P. R.....	1.0	0.9	0.8	2.0	2.2	2.4	0.8	0.8	0.8
	Sp. gr.	1.07	1.06	1.05	1.11	1.12	1.13	1.05	1.05	1.05
(3)	Evap.	20,600	15,800	12,650	20,100	16,800	14,600	21,900	16,800	13,400
	Per cent. solids..	18.5	14.0	12.0	13.0	15.0	16.5	14.0	11.5	10.7
	B. P. R.....	1.7	1.2	1.0	1.1	1.3	1.5	1.2	1.0	1.0
	Sp. gr.	1.10	1.08	1.07	1.07	1.08	1.09	1.08	1.08	1.06
(4)	Evap.	22,300	17,100	13,700	18,250	15,400	13,400	23,750	18,200	14,600
	Per cent. solids..	38.0	20.0	15.5	9.5	11.2	12.5	23.3	16.2	13.5
	B. P. R.....	5.0	1.8	1.4	0.8	1.0	1.1	2.3	1.4	1.1
	Sp. gr.	1.22	1.11	1.08	1.05	1.06	1.07	1.13	1.09	1.08
(5)	Evap.		18,400	14,700		14,000	12,300		19,600	15,700
	Per cent. solids..		38.0	21.5		9.0	10.5		26.5	17.5
	B. P. R.....		5.0	2.0		0.8	0.9		2.7	2.0
	Sp. gr.		1.22	1.12		1.05	1.06		1.14	1.10
(6)	Evap.			15,800			11,200			16,800
	Per cent. solids..			38.0			9.0			29.5
	B. P. R.....			5.0			0.7			3.4
	Sp. gr.			1.22			1.05			1.16

TEMPERATURE DISTRIBUTION AND STATIC HEAD LOSS.

PARALLEL CURRENT OPERATION (1).

Bodies	Quadruple		Quintuple		Sextuple	
	Temp.	Loss	Temp.	Loss	Temp.	Loss
Steam belt 1.....	240	..	240	..	240	..
Vapor belt 1.....	224	2.0	227	2.0	230	2.0
Vapor belt 2.....	205	2.0	214	2.0	220	2.0
Vapor belt 3.....	179	2.5	197	2.0	207	2.0
Vapor belt 4.....	115	11.0	173	3.0	190	2.3
Vapor belt 5.....	..	..	115	11.0	165	4.5
Vapor belt 6.....	..	..	..	..	115	11.0
Total static head L...		17.5		20.0		23.8

Taking these up one at a time, and beginning with the consideration of the quadruple effect, it is remembered that the total available temperature drop is 125° from the first steam belt to the condenser. The static head losses from the above table are 17.5° . The boiling point rise is given on the chart preceding, and amounts to a total of 8.5° . The total losses of temperature from these two sources are then 26° , leaving a net working temperature drop of 99° F. On the presumption of equal surfaces in all bodies of a given apparatus, which represents usual practice, especially where there has been no provision made for robbing vapors from the intermediate bodies for other purposes, it is now necessary to divide up the net temperature fall between the various bodies in proportion to the work each has to do as shown in the chart, and the conditions under which this work has to be done.

This is a most difficult task, and only approximate results can be expected on account of the practical impossibility of providing for all the variables. The coefficient of heat transmission, in the absence of actual data on the solution in the type of evaporator contemplated or a pilot thereof is a matter of guess work. It has been brought out before that this coefficient is affected by (1) the temperature of steam, (2) the viscosity of the solution, (3) the rate of evaporation, (4) the velocity of circulation, (5) the cleanliness of the heating surfaces, (6) the presence of non-condensable gases.

One must be satisfied with a compromise, under the circumstances, and an attempt will be made to establish the probable coefficients of heat transmission as determined by steam temperature and viscosity, neglecting the other factors.

By referring to the chapter on Heat Transmission, there will be found a diagram of Heat Transmissions for pure water with steam at varying temperatures, as established by Professor Kerr in his research work. The transmissions shown in the general trend curve will be used here. The temperature of the water boiling on the liquor side of the evaporators in these tests was roughly 20° F. lower than that of steam in the calandrias. Fig. 6 gives the viscosity of water at varying temperatures as well as the viscosity of sulphite liquor at various temperatures

DETERMINATION OF NET TEMPERATURE DROPS.

QUADRUPLE. CYCLE (1).

Body No.	Evapora- tion F.	Per Cent Solids	Liquor F.	Viscosity	Steam F.	Heat Trans. U.	Water Temp. for U.	Viscosity of Water	Cor. Fact.	For Liquor K.L.	Cor. Fact.	Ratio KL KW	Cor. Trans.	$\frac{F}{C}$	Temp. Drop
1	17,200	9.5	224.8	0.80	240	1,300	220	.75	1.18	1.22	1.22	.968	1,257	13.7	10.5
2	18,000	12.5	208.0	0.85	224	1,030	204	.79	1.15	1.19	1.19	.968	1,000	18.9	14.5
3	20,600	18.5	180.7	1.00	207	800	187	.82	1.05	1.17	1.17	.900	720	28.6	22.0
4	22,300	38.0	120.0	1.88	179	500	159	.83	0.68	1.15	1.15	.590	330	67.6	52.0
													Total		99.0

Quintuple. Cycle (1).

1	13,200	9.5	227.8	0.78	240	1,300	220	.75	1.20	1.22	1.22	.984	1,280	10.3	7.5
2	14,500	11.0	214.9	0.83	227	1,070	207	.78	1.16	1.20	1.20	.966	1,030	14.1	10.3
3	15,800	14.0	198.2	0.92	214	870	194	.80	1.10	1.18	1.18	.933	813	19.4	14.2
4	17,100	20.0	174.8	1.00	197	710	177	.83	1.05	1.16	1.16	.905	643	26.6	19.5
5	18,400	38.0	120.0	1.88	173	520	153	.84	0.68	1.15	1.15	.590	307	59.7	43.8
													Total		95.3

Sextuple. Cycle (1).

1	10,550	9.0	230.8	0.78	240	1,300	220	.75	1.20	1.22	1.22	.984	1,280	8.2	5.5
2	11,600	10.0	220.8	0.84	230	1,120	210	.76	1.16	1.21	1.21	.956	1,070	10.8	7.3
3	12,650	12.0	208.0	0.90	220	970	200	.77	1.11	1.20	1.20	.925	897	14.1	9.5
4	13,700	15.5	191.4	0.96	207	810	187	.82	1.07	1.17	1.17	.915	740	18.5	12.5
5	14,700	21.5	167.0	1.00	190	640	170	.83	1.05	1.16	1.16	.905	578	25.4	17.1
	15,800	38.0	120.0	1.88	165	470	145	.85	0.68	1.15	1.15	.590	277	57.0	38.3
													Total		90.2

and densities for the ranges encountered in the evaporators under discussion.

Then by reference to Figs. 5 and 6, correction factors can be determined for the selected transmissions as determined by the temperature existing in the steam belts, and corresponding to the viscosities existing. In the table above, all of this data has been systematized and does not need very much elaboration. U is the heat transmission when evaporating water, taken from Professor Kerr. U_c is the same coefficient corrected for the relative viscosity in proportion to the correction factors for water and liquor as shown in the diagrams referred to above. The total temperature difference, net, is distributed in proportion to the evaporation in each body, and inversely proportional to the corrected heat transmission, and this distribution is given in the last column. The same data have been worked out for the quintuples and sextuples.

The corrected distribution for the temperatures and latent heats are given below for the various units of cycle I.

DISTRIBUTION FOR QUADRUPLE EFFECT (1).

Bodies	S.B.	1	2	3	4	Total
B. P. R.....		0.8	1.0	1.7	5.0	8.5
S. H. loss.....		2.0	2.0	2.5	11.0	17.5
Net drop		10.5	14.5	22.0	52.0	99.0
Total drop		13.3	17.5	26.2	68.0	125.0
Round fig.		13.0	18.0	26.0	68.0	125.0
Temp. vap.	240.0	227.0	209.0	183.0	115.0	
Latent heat	952.1	960.7	972.2	988.1	1,027.2	
Boiling points		227.8	210.0	184.7	120.0	
Round fig.		228.0	210.0	185.0	120.0	

DISTRIBUTION FOR QUINTUPLE EFFECT (1).

Bodies	S.B.	1	2	3	4	5	Total
B. P. R.....		0.8	0.9	1.2	1.8	5.0	9.7
S. H. loss.....		2.0	2.0	2.0	3.0	11.0	20.0
Net drop		7.5	10.3	14.2	19.5	43.8	95.3
Total drop		10.3	13.2	17.4	24.3	59.8	125.0
Round fig.		10.0	13.0	18.0	24.0	60.0	125.0
Temp. vap.	240.0	230.0	217.0	199.0	175.0	115.0	
Latent heat	952.1	958.7	967.2	978.4	992.9	1,027.2	
Boil. temp.		230.8	217.9	202.2	176.8	120.0	
Round fig.		231.0	218.0	202.0	177.0	120.0	

DISTRIBUTION FOR SEXTUPLE EFFECT (1).

Bodies	S.B.	1	2	3	4	5	6
B. P. R.....		0.8	0.8	1.0	1.4	2.0	5.0
S. H. loss.....		2.0	2.0	2.0	2.3	4.5	11.0
Net drop		5.5	7.3	9.5	12.5	17.1	38.3
Total drop		8.3	10.1	12.5	16.2	23.6	54.3
Round fig.		8.0	10.0	13.0	16.0	24.0	54.0
Temp. vap.	240.0	232.0	222.0	209.0	193.0	169.0	115.0
Latent heat	952.1	957.4	963.9	972.2	982.1	996.4	1,027.2
Boil. temp.		232.8	222.8	210.0	194.4	171.0	120.0
Round fig.		233.0	223.0	210.0	194.0	171.0	120.0

The distribution of temperatures in the quadruples, quintuples, and sextuples operating counter current as per cycle 2 carrying the concentration to the ultimate of 50 per cent. solids is determined in a similar manner to the above, and the probable distribution of evaporation is given in the tentative list as before, on page 366, as well as the boiling point rise, B.P.R. As in the former case, it now becomes necessary to make a provisional distribution of temperature fall through the various bodies. In this case, it must be remembered that the evaporation is maximum in the first body, and also the concentration of liquor. It is therefore to be expected that this distribution will be quite different from what it was before. This is given below approximately, as well as the calculated static head losses.

TEMPERATURE DISTRIBUTION AND STATIC HEAD LOSS.

COUNTER CURRENT OPERATION (2).

Bodies	Quadruple		Quintuple		Sextuple	
	Temp.	Loss	Temp.	Loss	Temp.	Loss
Steam belt 1.....	240	..	240	..	240	..
Vapor belt 1.....	197	2.0	203	2.0	207	2.0
Vapor belt 2.....	178	2.6	188	2.3	193	2.2
Vapor belt 3.....	157	3.7	171	3.2	181	2.6
Vapor belt 4.....	115	9.7	152	4.8	167	3.2
Vapor belt 5.....	..	..	115	9.7	148	5.0
Vapor belt 6.....	..	..	..	..	115	9.7
Total losses		18.0		22.0		24.7

The temperature distribution is arrived at in the same manner as before, and similar data are given in the determination.

TEMPERATURE DISTRIBUTION QUADRUPLE COUNTER CURRENT (2).

Bodies	S.B.	1	2	3	4
B. P. R.....		9.5	2.0	1.1	0.8
S. H. loss.....		2.0	2.6	3.7	9.7
Net drop		24.6	21.0	22.4	25.6
Total drop		36.1	25.6	27.2	36.1
Round fig.		36.0	26.0	27.0	36.0
Temp. vap.	240.0	204.0	178.0	151.0	115.0
Latent heat	952.1	975.4	991.1	1,006.8	1,027.2
Boil. temp.		213.5	180.0	152.1	115.8
Round fig.		214.0	180.0	152.0	116.0

TEMPERATURE DISTRIBUTION QUINTUPLE COUNTER CURRENT (2).

Bodies	S.B.	1	2	3	4	5
B. P. R.....		9.5	2.2	1.3	1.0	0.8
S. H. loss.....		2.0	2.3	3.2	4.8	9.7
Net drop		18.7	15.1	16.7	18.0	10.7
Total drop		30.2	19.6	21.2	23.8	30.2
Round fig.		30.0	20.0	21.0	24.0	30.0
Temp. vap.	240.0	210.0	190.0	169.0	145.0	115.0
Latent heat	952.1	971.6	983.9	996.4	1,010.3	1,027.2
Boil. temp.		219.5	192.2	170.3	146.0	115.8
Round fig.		220.0	192.0	170.0	146.0	116.0

TEMPERATURE DISTRIBUTION SEXTUPLE, COUNTER CURRENT (2).

Bodies	S.B.	1	2	3	4	5	6
B. P. R.....		9.5	2.4	1.5	1.1	0.9	0.8
S. H. loss.....		2.0	2.2	2.6	3.2	5.0	9.7
Net drop		15.4	11.9	12.8	13.8	14.0	15.4
Total drop		26.9	16.5	16.9	18.1	20.8	25.9
Round fig.		27.0	16.0	17.0	18.0	21.0	26.0
Temp. vap.	240.0	213.0	197.0	180.0	162.0	141.0	115.0
Latent heat	952.1	969.7	979.7	989.9	1,000.5	1,012.6	1,027.2
Boil. temp.		222.5	199.4	181.5	163.1	141.9	115.8
Round fig.		223.0	199.0	182.0	163.0	142.0	116.0

As explained before, it is not claimed that these distributions are absolutely correct, but they are close enough to be used in making up the heat balances which follow with reasonable accuracy.

As regards the operation of multiple effects according to cycle (3), the distribution of temperatures is not very different from that in cycle (2). The data are developed by the same process of reasoning as before, and are shown below.

TENTATIVE DISTRIBUTION OF TEMPERATURES AND STATIC HEAD LOSS.

CYCLE (3), FEED SEQUENCE, SECOND, THIRD, FOURTH, FIFTH, FIRST.

Bodies	Quadruple		Quintuple		Sextuple	
	Temp.	Loss	Temp.	Loss	Temp.	Loss
Steam belt 1.....	240	..	240	..	240	..
Vapor belt 1.....	205	2.0	210	2.0	214	2.0
Vapor belt 2.....	189	2.0	198	2.0	205	2.0
Vapor belt 3.....	164	3.0	182	2.6	193	2.2
Vapor belt 4.....	115	10.0	160	4.0	178	2.5
Vapor belt 5.....	..	..	115	10.0	157	4.4
Vapor belt 6.....	..	..	..	..	115	10.0
Total static head losses		17.0		20.6		23.1

The log below shows the determination of the distribution of the net temperature drop, as before.

The final distribution of temperature drops for cycle 3 is given below for the various sets.

TEMPERATURE DISTRIBUTION FOR QUADRUPLE CYCLE (3).

Bodies	S.B.	1	2	3	4
B. P. R.....		9.5	0.8	1.2	2.3
S. H. loss.....		2.0	2.0	3.0	10.0
Net drop		18.7	15.7	22.1	37.7
Total drop		30.2	18.5	26.3	50.0
Round fig.		30.0	18.0	27.0	50.0
Temp. vap.	240.0	210.0	192.0	165.0	115.0
Latent heat	952.1	971.6	982.7	998.7	1,027.2
Boil. temp.		219.5	192.8	166.2	117.3
Round fig.		220.0	193.0	166.0	117.0

DETERMINATION OF NET TEMPERATURE DROP.

QUADRUPEL. CYCLE (3).

Body No.	Evapora- tion F.	Per Cent Solids	Liquor Temp. F.	Viscosity	Steam F.	Heat Trans. U.	Water Temp. for U.	Viscosity of Water	Cor. Fact. for Water	Cor. Fact. for Liquor	Ratio KL KW	U _c Cor. Trans.	$\frac{U_c}{U}$	Temp. Drop
Quintuple.														
1	14,000	50.0	210.5	2.65	240	1,300	220	.76	1.21	0.55	.454	591	30.9	18.7
2	15,400	9.5	108.8	1.00	210	850	190	.79	1.19	1.17	.983	776	25.9	15.7
3	16,800	11.5	183.0	0.93	198	710	178	.81	1.17	1.12	.958	680	24.7	14.1
4	18,200	16.2	161.9	0.94	182	570	162	.83	1.16	1.05	.915	521	35.0	19.9
5	19,600	26.5	117.7	0.98	160	450	140	.87	1.13	0.90	.796	358	54.8	31.2
												Total		89.0
Sextuple.														
1	11,200	50.0	223.5	2.60	240	1,300	220	.76	1.21	0.57	.472	613	18.3	10.2
2	12,300	9.0	205.8	1.00	214	900	194	.78	1.20	1.18	.983	885	13.9	7.6
3	13,400	10.7	194.0	0.94	205	790	185	.80	1.18	1.14	.965	762	17.6	9.6
4	14,600	13.5	179.1	0.92	193	670	173	.82	1.17	1.10	.940	630	23.2	12.7
5	15,700	17.5	159.0	0.96	178	550	158	.84	1.16	1.05	.906	498	31.5	17.3
6	16,800	29.0	118.4	0.98	157	440	137	.88	1.12	0.88	.785	345	48.7	26.7
												Total		84.1

TEMPERATURE DISTRIBUTION FOR QUINTUPLE CYCLE (3).

Bodies	S.B.	1	2	3	4	5
B. P. R.....		9.5	0.8	1.0	1.4	2.7
S. H. loss		2.0	2.0	2.6	4.0	10.0
Net drop		13.3	10.5	14.1	19.9	31.2
Total drop		24.8	13.3	17.7	25.3	43.9
Round fig.		25.0	13.0	18.0	25.0	44.0
Temp. vap.	240.0	215.0	202.0	184.0	159.0	115.0
Latent heat.	952.1	968.4	976.6	987.5	1,002.2	1,027.2
Boil. temp.		224.5	202.8	185.0	160.4	117.7
Round fig.		225.0	203.0	185.0	160.0	118.0

TEMPERATURE DISTRIBUTION FOR SEXTUPLE, CYCLE (3).

Bodies	S.B.	1	2	3	4	5	6
B. P. R.....		9.5	0.8	1.0	1.1	2.0	3.4
S. H. loss		2.0	2.0	2.2	2.5	4.4	10.0
Net drop		10.2	7.6	9.6	12.7	17.3	26.7
Total drop		21.7	10.4	12.8	16.3	23.7	40.1
Round fig.		22.0	10.0	13.0	16.0	24.0	40.0
Temp. vap.	240.0	218.0	208.0	195.0	179.0	155.0	115.0
Latent heat	952.1	966.5	972.9	980.9	990.5	1,004.5	1,027.2
Boil. temp.		227.5	208.8	196.0	180.1	157.0	118.4
Round fig.		228.0	209.0	196.0	180.0	157.0	118.0

This completes the distribution of temperature drops for the various multiple effects under discussion operating according to the different cycles suggested. This data will now be used in making up the heat balances showing the work of each apparatus.

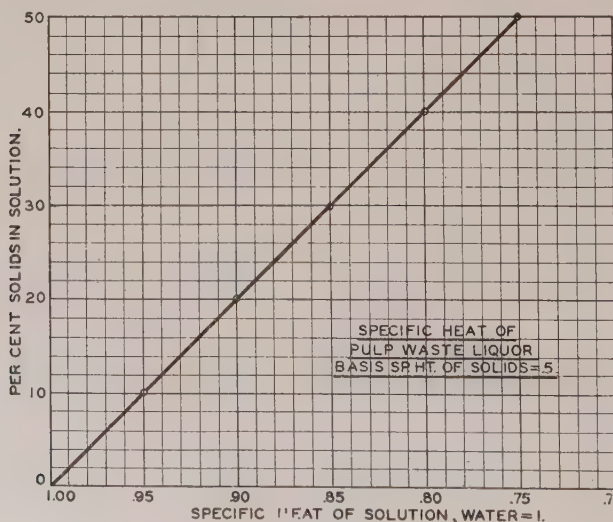


FIG. 7.

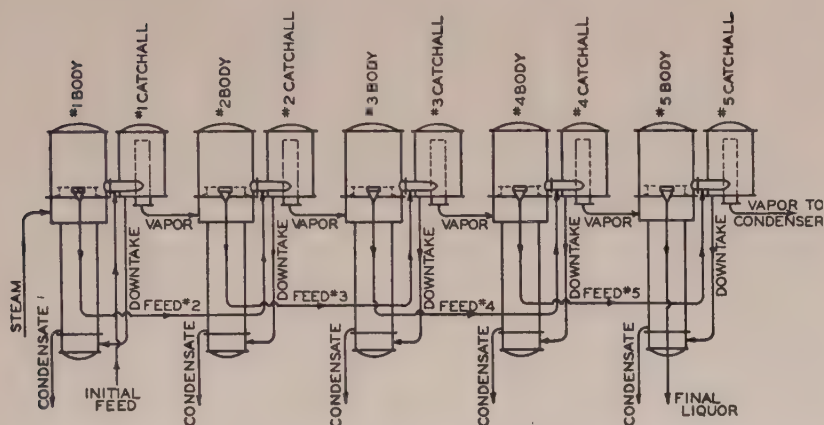


FIG. 8.—Diagrammatic Sketch of Quintuple Effect Evaporator for Evaporating Paper Mill Waste Liquors.

Cycle No. 1, Parallel Current—Feed, Sequence 1-2-3-4-5.

In order to make full allowance for the variables in making up the heat balances the specific heat of the solution at various concentrations must be taken into account. This is a refinement that is not absolutely necessary, but it is incorporated in the calculations that follow. Fig. 7

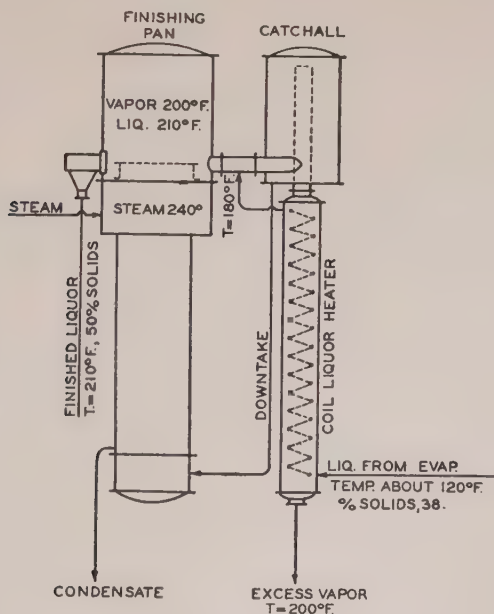


FIG. 9.—Finishing Pan for Paper Mill Waste Liquor. Cycle No. 1.

shows the relation between the specific heat of solution, and the percentage solids in solution. Whenever there will be a flash or a heating operation, therefore, this correction will be made, without further explanation.

Fig. 8 shows a diagrammatic arrangement of the quintuple effect operating according to cycle 1.

The work of the multiple effects in this cycle is based upon an evaporation of 79,000 lbs. per hour, the balance, 5,000 lbs. per hour, being done in a separate finishing pan as shown in Fig. 9, all as explained in the discussion preceding. Before entering the finisher, the liquor from the multiple effect is preheated by means of a coil heater installed as part of the vapor pipe leaving the pan, thus giving a large excess of vapors and good steam circulation. The coil is restricted in area for the purpose of maintaining a high liquor velocity while the heating operation is going on, thus securing good transmission. Under such conditions it is estimated that it will be easily possible to bring the temperature of the liquor to within 20° F. of the temperature of the vapors. The assumed conditions of operation of the finisher are shown in the sketch.

The attached heat balances show in detail the work of the quadruples, quintuples, and sextuples operating according to cycle 1. A heat balance of the finisher and heater is also given.

In view of the elaboration given as to the methods used in figuring these heat balances in the chapter on The Heat Balance, this information is plain enough, and hardly needs further explanation.

HEAT BALANCE, QUADRUPLE, CYCLE (1), SEQUENCE, 1-2-3-4.

Body	Specifications	Heat	Liquor
(1) Steam, 23,350 lbs. @ 240 =	$23,350 \times 952.1 = \dots\dots\dots$	22,200,000	100,000
Ded. for heat, 100,000 $\times (228 - 175 = 53) \times 0.96 = \dots$		<u>5,090,000</u>	
Available for evaporation.....		17,110,000	
L @ 227 = 960.7. $E_1 = \frac{17,110,000}{960.7} = \dots\dots\dots$			<u>17,800</u>
Transferred to No. 2.....			82,200
(2) Vapor ex. No. 1.....		17,110,000	
Add flash, 82,200 $\times (228 - 210 = 18) \times 0.95 = \dots\dots$		<u>1,405,000</u>	
Available for evaporation.....		18,515,000	
L @ 209 = 972.2. $E_2 = \frac{18,515,000}{972.2} = \dots\dots\dots$			<u>19,050</u>
Transferred to No. 3.....			63,150
(3) Vapor ex. No. 2.....		18,515,000	
Add flash, 63,150 $\times (210 - 185 = 25) \times 0.935 = \dots\dots$		<u>1,480,000</u>	
Available for evaporation.....		19,995,000	
L @ 183 = 988.1. $E_3 = \frac{19,995,000}{988.1} = \dots\dots\dots$			<u>20,250</u>
Transferred to No. 4.....			42,900
(4) Vapor ex. No. 3.....		19,995,000	
Add flash, 42,900 $\times (185 - 120 = 65) \times 0.92 = \dots\dots$		<u>2,560,000</u>	
Available for evaporation.....		22,555,000	
L @ 115 = 1027.2. $E_4 = \frac{22,555,000}{1,027.2} = \dots\dots\dots$			<u>21,900</u>
Concentrate ex. No. 4.....			21,000

HEAT BALANCE, QUINTUPLE, CYCLE (1), SEQUENCE, 1-2-3-4-5.

Body	Specifications	Heat	Liquor
(1)	Steam, 19,400 lbs. @ $240 = 19,400 \times 952.1 = \dots\dots\dots$	18,470,000	100,000
	Ded. for heat. $100,000 \times (231 - 175 = 56) \times 0.96 = \dots\dots\dots$	<u>5,370,000</u>	
	Available for evaporation.....	13,100,000	
	L @ 230 = 958.7. $E_1 = \frac{13,100,000}{958.7} = \dots\dots\dots$		<u>13,650</u>
	Transferred to No. 2.....		86,350
(2)	Vapor ex. No. 1.....	13,100,000	
	Add flash, $86,350 \times (231 - 218 = 13) \times 0.952 = \dots\dots\dots$	<u>1,167,000</u>	
	Available for evaporation.....	14,267,000	
	L @ 217 = 967.2. $E_2 = \frac{14,267,000}{967.2} = \dots\dots\dots$		<u>14,750</u>
	Transferred to No. 3.....		71,600
(3)	Vapor ex. No. 2.....	14,267,000	
	Add flash, $71,600 \times (218 - 200 = 18) \times 0.945 = \dots\dots\dots$	<u>1,218,000</u>	
	Available for evaporation.....	15,485,000	
	L @ 199 = 978.4. $E_3 = \frac{15,485,000}{978.4} = \dots\dots\dots$		<u>15,800</u>
	Transferred to No. 4.....		55,800
(4)	Vapor ex. No. 3.....	15,485,000	
	Add flash, $55,800 \times (200 - 177 = 23) \times 0.93 = \dots\dots\dots$	<u>1,194,000</u>	
	Available for evaporation.....	16,679,000	
	L @ 175 = 992.9. $E_4 = \frac{16,679,000}{992.9} = \dots\dots\dots$		<u>16,700</u>
	Transferred to No. 5.....		39,100
(5)	Vapor ex. No. 4.....	16,679,000	
	Add flash, $39,100 \times (177 - 120 = 57) \times 0.90 = \dots\dots\dots$	<u>2,005,000</u>	
	Available for evaporation.....	18,684,000	
	L @ 115 = 1027.2. $E_5 = \frac{18,684,000}{1,027.2} = \dots\dots\dots$		<u>18,150</u>
	Concentrate ex. No. 5.....		20,950

HEAT BALANCE, SEXTUPLE, CYCLE (1), SEQUENCE, 1-2-3-4-5-6.

Body	Specifications	Heat	Liquor
(1)	Steam, 16,900 lbs. @ $240 = 16,900 \times 952.1 = \dots\dots\dots$	16,100,000	100,000
	Deduct for heating, $100,000 \times (233 - 175 = 58) \times 0.96 = \dots\dots\dots$	<u>5,565,000</u>	
	Available for evaporation.....	10,535,000	
	L @ 232 = 957.4. $E_1 = \frac{10,535,000}{957.4} = \dots\dots\dots$		<u>11,000</u>
	Transferred to No. 2.....		89,000
(2)	Vapor ex. No. 1.....	10,535,000	
	Add flash, $89,000 \times (233 - 223 = 10) \times 0.955 = \dots\dots\dots$	<u>850,000</u>	
	Available for evaporation.....	11,385,000	
	L @ 222 = 963.9. $E_2 = \frac{11,385,000}{963.9} = \dots\dots\dots$		<u>11,800</u>
	Transferred to No. 3.....		77,200

HEAT BALANCE, SEXTUPLE, CYCLE (1), SEQUENCE, 1-2-3-4-5-6.

Body	Specifications	Heat	Liquor
(3)	Vapor ex. No. 2.....	11,385,000	
	Add flash, $77,200 \times (223 - 210 = 13) \times 0.95 =$	<u>953,000</u>	
	Available for evaporation.....	12,338,000	
	L @ 209 = 972.2. $E_3 = \frac{12,338,000}{972.2} =$		12,700
	Transferred to No. 4.....		64,500
(4)	Vapor ex. No. 3.....	12,338,000	
	Add flash, $64,500 \times (210 - 194 = 16) \times 0.94 =$	<u>970,000</u>	
	Available for evaporation.....	13,308,000	
	L @ 193 = 982.1. $E_4 = \frac{13,308,000}{982.1} =$		13,550
	Transferred to No. 5.....		50,950
(5)	Vapor ex. No. 4.....	13,308,000	
	Add flash, $59,950 \times (194 - 171 = 23) \times 0.92 =$	<u>1,079,000</u>	
	Available for evaporation.....	14,387,000	
	L @ 169 = 996.4. $E_5 = \frac{14,387,000}{996.4} =$		14,450
	Transferred to No. 6.....		36,500
(6)	Vapor ex. No. 5.....	14,387,000	
	Add flash, $36,500 \times (171 - 120 = 51) \times 0.89 =$	<u>1,660,000</u>	
	Available for evaporation.....	16,047,000	
	L @ 115 = 1027.2. $E_6 = \frac{16,047,000}{1,027.2} =$		15,600
	Concentrate ex. No. 6.....		20,900

HEAT BALANCE FINISHING PAN AND HEATER.

CYCLE (1) PARALLEL CURRENT.

Operation, evaporate 5,000 lbs. of water from 21,000 lbs. of liquor containing 8,000 lbs. of solids, and at a temperature of 120° F. Steam temperature, 240° F.; vapor temperature, 200° F. Vapor heat liquor before entering evaporator from 120° F. to 180° F.

Heat in vapors ex. finisher, 5,000 lbs. @ 200° F. = $5,000 \times 977.8 =$	4,880,000 B.t.u.'s	
Heat liquor 180° F. to 210° F., $21,000 \times (210 - 180 = 30) \times 0.81 =$	<u>510,000 B.t.u.'s</u>	
Heat requirements steam belt.....	5,390,000 B.t.u.'s	
L @ 240 = 952.1. Steam to finisher = $\frac{5,390,000}{952.1} =$		5,670 lbs.
For vapor heater, Vapors ex. finisher		5,000 lbs.
Heater requirements, $21,000 \times (180 - 120 = 60) \times 0.81 =$	1,020,000 B.t.u.'s	
L @ 200 = 977.8. Vapors corresponding, $\frac{1,020,000}{977.8} =$..		1,044 lbs.
Excess vapors		3,956 lbs.

It is well to point out that there are still a number of things that can be done to improve the economy of these cycles, details of which are omitted on account of the complication of figures involved. The following are pointed out for the interest of those concerned.

1. The excess vapors from the finishing pan can be put into the steam belt of the third effect, thus utilizing this vapor in more stages, according to the number of bodies, instead of destroying them. There is really no undesirable complication involved.

2. The liquor comes to the multiple effect at a temperature of 175° F. It has been brought out in the chapter on Vapor Heating, and more

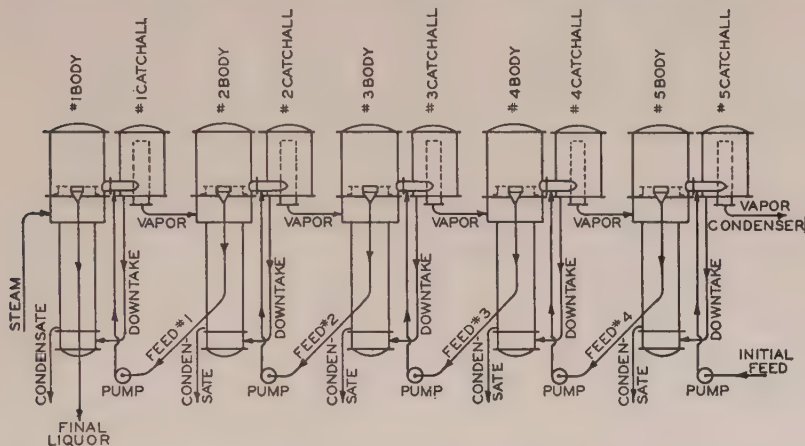


FIG. 10.—Diagrammatic Sketch of Quintuple Effect Evaporator for Evaporating Paper Mill Waste Liquors.

Cycle No. 2, Sequence of Feed, 5-4-3-2-1.

elaborately in the chapter on the Heat Balance in the Beet Sugar Factory, that vapors can be used to great advantage to raise the temperature of solutions, and that one can heat in this way to within 10° of the temperature of the vapor used. If, therefore, the amount of heating possible in this way amounts to a substantial quantity, this principle can be used advantageously. Cycle 1 assumes that the liquor is fed directly into the first body, and remembering the temperature distribution, we can accomplish the following:

In the quadruple, liquor can be heated from 175° F. to 199° F. with No. 2 vapors, and from 199° F. to 217° F. with No. 1 vapors.

In the quintuple, liquor can be heated from 175° F. to 189° F. with No. 3 vapors, to 207° F. with No. 2 vapors, and 220° F. with No. 1 vapors.

In the sextuple, liquor can be heated from 175° F. to 199° F. with No. 3 vapors, to 212° F. with No. 2 vapors, and 222° F. with No. 1 vapors.

3. The economy of this cycle can also be improved by by-passing the condensate, or its flash, from each steam belt to the next, reducing the steam consumption perhaps 4 per cent., depending upon the operating conditions. This also was elaborated in the text in the chapter on Example of the Use of the Heat Balance.

Fig. 10 shows a diagrammatic arrangement of a quintuple effect operating as per cycle 2.

Here, as was explained before, the entire work of evaporation is done in the multiple effect which must evaporate 84,000 lbs. of water per hour from the 100,000 lbs. per hour of liquor, thus raising its density from 8 per cent. solids to 50 per cent. solids. The initial temperature of the liquor as it reaches the apparatus is taken at 175° as in all cases. Detailed explanations of the method of figuring the heat balances will be found in the same reference as before.

The balance sheets are all shown in detail showing the contemplated performance, and need no further elaboration.

HEAT BALANCE, QUADRUPLE, CYCLE (2), SEQUENCE 4-3-2-1.

Body	Specifications	Heat	Liquor
(4)	Heat ex. No. 4, 22,950 lbs. @ 115 = $22,950 \times 1,027.2 =$	23,550,000	100,000
	Ded. flash, 100,000 $\times (175 - 116 = 59) \times 0.96 = \dots$	5,660,000	22,950
	Heat requirements S.B. 4 =	17,890,000	77,050
	L @ 151 = 1006.8. $E_3 = \frac{17,890,000}{1,006.8} = \dots$		17,750
	Transferred to No. 2.....		59,300
(3)	Vapor to No. 4.....	17,890,000	
	Add heat. 77,050 $\times (152 - 116 = 36) \times 0.95 = \dots$	2,635,000	
	Heat requirements S.B. 3 =	20,525,000	
	L @ 178 = 991.1. $E_2 = \frac{20,525,000}{991.1} = \dots$		20,700
	Transferred to No. 1.....		38,600
(2)	Vapor to No. 3.....	20,525,000	
	Add heat. 59,300 $\times (180 - 152 = 28) \times 0.93 = \dots$	1,552,000	
	Heat requirements S.B. 2 =	22,077,000	
	L @ 204 = 975.4. $E_1 = \frac{22,077,000}{975.4} = \dots$		22,600
	Concentrate ex. No. 1.....		16,000
(1)	Vapor to No. 2.....	22,077,000	
	Add heat. 38,600 $\times (214 - 180 = 34) \times 0.895 = \dots$	1,172,000	
	Heat requirements S.B. 1.....	23,249,000	
	L @ 240 = 952.1. Steam to No. 1 = $\frac{23,249,000}{952.1} = \dots$		24,400

HEAT BALANCE, QUINTUPLE, CYCLE (2), SEQUENCE 5-4-3-2-1.

Body	Specifications	Heat	Liquor
(5)	Heat ex. No. 5, 18,550 lbs. @ 115 = $18,550 \times 1,027.2 =$	19,050,000	100,000
	Ded. flash, $100,000 \times (175 - 116 = 59) \times 0.96 = \dots$	5,660,000	E5 = 18,550
	Heat requirements S.B. 5 =	13,390,000	81,450
	L @ 145 = 1010.3. E4 = $\frac{13,390,000}{1,010.3} = \dots$		13,230
	Transferred to No. 3.		68,220
(4)	Vapor to No. 5.	13,390,000	
	Add heat, $81,450 \times (146 - 116 = 30) \times 0.95 = \dots$	2,320,000	
	Heat requirements S.B. 4.	15,710,000	
	L @ 169 = 996.4. E3 = $\frac{15,710,000}{996.4} = \dots$		15,760
	Transferred to No. 2.		52,460
(3)	Vapor to No. 4.	15,710,000	
	Add heat, $68,220 \times (170 - 146 = 24) \times 0.94 = \dots$	1,540,000	
	Heat requirements S.B. 3.	17,250,000	
	L @ 190 = 983.9. E2 = $\frac{17,250,000}{983.9} = \dots$		17,550
	Transferred to No. 1.		34,910
(2)	Vapor to No. 3.	17,250,000	
	Add heat, $52,460 \times (192 - 170 = 22) \times 0.915 = \dots$	1,054,000	
	Heat requirements S.B. 2.	18,304,000	
	L @ 210 = 971.6. E1 = $\frac{18,304,000}{971.6} = \dots$		18,900
	Concentrate ex. No. 1.		16,010
(1)	Vapor to No. 2.	18,304,000	
	Add heat, $34,910 \times (220 - 192 = 28) \times 0.885 = \dots$	865,000	
	Heat requirements S.B. 1.	19,169,000	
	L @ 240 = 952.1. Steam to No. 1 = $\frac{19,169,000}{952.1} = \dots$		20,100

HEAT BALANCE, SEXTUPLE, CYCLE (2), SEQUENCE, 6-5-4-3-2-1.

Body	Specifications	Heat	Liquor
(6)	Heat ex. No. 6, 15,600 lbs. @ 115 = $15,600 \times 1,027.2 =$	16,030,000	100,000
	Ded. flash, $100,000 \times (175 - 116 = 59) \times 0.96 = \dots$	5,660,000	E6 = 15,600
	Heat requirements S.B. 6.	10,370,000	84,400
	L @ 141 = 1,012.6. E5 = $\frac{10,370,000}{1,012.6} = \dots$		10,250
	Transferred to No. 4.		74,150

HEAT BALANCE, SEXTUPLE, CYCLE (2), SEQUENCE, 6-5-4-3-2-1—
 (Continued).

Body	Specifications	Heat	Liquor
(5)	Vapor to No. 6.....	10,370,000	
	Add heat. $84,400 \times (142 - 116 = 26) \times 0.952 =$	2,090,000	
	Heat requirements S.B. 5.....	12,460,000	
	L @ 162 = 1000.5. $E_4 = \frac{12,460,000}{1,000.5} =$		12,460
	Transferred to No. 3.....		61,690
(4)	Vapor to No. 5.....	12,460,000	
	Add heat. $74,150 \times (163 - 142 = 21) \times 0.945 =$	1,470,000	
	Heat requirements S.B. 4.....	13,930,000	
	L @ 180 = 989.9. $E_3 = \frac{13,930,000}{989.9} =$		14,090
	Transferred to No. 2.....		47,600
(3)	Vapor to No. 4.....	13,930,000	
	Add heat $61,690 \times (182 - 163 = 19) \times 0.935 =$	1,093,000	
	Heat requirements S.B. 3.....	15,023,000	
	L @ 197 = 979.7. $E_2 = \frac{15,023,000}{979.7} =$		15,350
	Transferred to No. 1.....		32,250
(2)	Vapor to No. 3.....	15,023,000	
	Add heat $47,600 \times (199 - 182 = 17) \times 0.915 =$	739,000	
	Heat requirements S.B. 2.....	15,762,000	
	L @ 213 = 969.7. $E_1 = \frac{15,762,000}{969.7} =$		16,250
	Concentrate ex. No. 1.....		16,000
(1)	Vapor to No. 2.....	15,762,000	
	Add heat $32,250 \times (223 - 199 = 24) \times 0.875 =$	677,000	
	Heat requirements S.B. 1.....	16,439,000	
	L @ 240 = 952.1. Steam to No. 1 = $\frac{16,439,000}{952.1} =$		17,270

As in the former case, it is possible to improve materially the performance as regards heat economy as follows, the detailed calculations having been left out, being too voluminous.

1. Instead of the liquor going directly into the last body, it can be made to "flash" down in successive flash pots, the vapors from each step going to a corresponding steam belt, and thus this flash can do useful work of evaporation in the latter bodies of the evaporator, instead of being totally destroyed in the condenser.

For the quadruple, flash down from 175° F. to 151° F. into the steam belt of the fourth effect, thus releasing about 2,300,000 B.t.u.'s with corresponding evaporation.

For the quintuple effect, flash down into the fourth steam belt from 175° to 169° releasing about 575,000 B.t.u.'s available for two stages of evaporation; and from 169° to 145° in the fifth steam belt thus releasing about 2,210,000 B.t.u.'s available for one stage of evaporation.

For the sextuple effect, flash down into the fifth steam belt from 175°

to 162° releasing about 1,250,000 B.t.u.'s available for two stages of evaporation; and from 162° to 141° into the steam belt of the sixth body releasing about 2,020,000 B.t.u.'s available for one stage of evaporation.

2. The feed from each body to the previous one can be heated through a coil heater with vapors leaving the body into which it is entering. The range is so small, however, that it is doubtful if the gain justifies the complication.

3. The condensate from the steam belts can be by-passed as in the case of cycle 1 with corresponding economy.

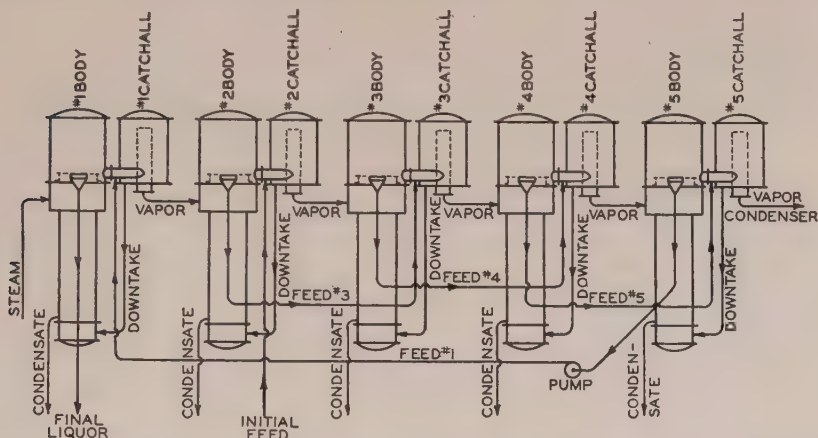


FIG. 11.—Diagrammatic Sketch of Quintuple Effect Evaporator for Evaporating Paper Mill Waste Liquors.

Cycle No. 3, Sequence of Feed, 2-3-4-5-1.

Fig. 11 shows a diagrammatic arrangement of a quintuple effect operating as per cycle 3.

The load is the same as for the other combinations. The method of calculating the heat balances submitted is by the trial and error, beginning by assuming the evaporation in the second body, and proceeding from this point until the total balances to the desired quantity of 84,000 lbs. per hour from the original of 100,000 lbs. of initial feed as before. In this way it is very similar to the parallel current principle as used in cycle 1.

The heat balances follow:

HEAT BALANCE, QUADRUPLE, CYCLE (3), SEQUENCE 2-3-4-1.

Body	Specifications	Heat	Liquor
(2)	Vapor ex. No. 1, 21,100 lbs. @ 210 = $21,100 \times 971.6 =$	20,500,000	100,000
	Ded. for heating 100,000 $\times (193 - 175 = 18) \times 0.96 =$	1,727,000	
	Available for evaporation.....	18,773,000	
	L @ 192 = 982.7. $E_2 = \frac{18,773,000}{982.7} =$		19,100
	Transferred to No. 3.....		80,900

HEAT BALANCE, QUADRUPLE, CYCLE (3), SEQUENCE 2-3-4-1
 (Continued).

Body	Specifications	Heat	Liquor
(3)	Vapor ex. No. 2.....	18,773,000	
	Add flash, $80,900 \times (193 - 166 = 27) \times 0.95 =$	2,075,000	
	Available for evaporation.....	20,848,000	
	L @ 165 = 998.7. $E_3 = \frac{20,848,000}{998.7} =$		20,850
	Transferred to No. 4.....		60,050
(4)	Vapor ex. No. 3.....	20,848,000	
	Add flash, $60,050 \times (166 - 117 = 49) \times 0.935 =$	2,750,000	
	Available for evaporation.....	23,598,000	
	L @ 115 = 1027.2. $E_4 = \frac{23,598,000}{1027.2} =$		22,950
	Transferred to No. 1.....		37,100
	Evaporation No. 1 (assumed).....		21,100
	Concentrate ex. No. 1.....		16,000
(1)	Evaporation by assumption, $21,100 \times 971.6 =$	20,500,000	
	Add heat, $37,100 \times (220 - 117 = 103) \times 0.892 =$	3,410,000	
	Heat requirements S.B. 1.....	23,910,000	
	L @ 240 = 952.1. Steam to No. 1 = $\frac{23,910,000}{952.1} =$		25,100

HEAT BALANCE, QUINTUPLE, CYCLE (3), SEQUENCE 2-3-4-5-1.

Body	Specifications	Heat	Liquor
(2)	Vapor ex. No. 1, 17,500 lbs. @ 215 = $17,500 \times 968.4 =$	16,900,000	100,000
	Ded. for heating $100,000 \times (203 - 175 = 28) \times 0.96 =$	2,690,000	
	Available for evaporation.....	14,210,000	
	L @ 202 = 976.6. $E_2 = \frac{14,210,000}{976.6} =$		14,560
	Transferred to No. 3.....		85,440
(3)	Vapor ex. No. 2.....	14,210,000	
	Add flash, $85,440 \times (203 - 185 = 18) \times 0.952 =$	1,445,000	
	Available for evaporation.....	15,655,000	
	L @ 184 = 987.5. $E_3 = \frac{15,655,000}{987.5} =$		15,870
	Transferred to No. 4.....		69,570
(4)	Vapor ex. No. 3.....	15,655,000	
	Add flash, $69,570 \times (185 - 160 = 25) \times 0.942 =$	1,636,000	
	Available for evaporation.....	17,291,000	
	L @ 159 = 1002.2. $E_4 = \frac{17,291,000}{1002.2} =$		17,270
	Transferred to No. 5.....		52,300

HEAT BALANCE, QUINTUPLE, CYCLE (3), SEQUENCE, 2-3-4-5-1
(Continued)

Body	Specifications	Heat	Liquor
(5)	Vapor ex. No. 4.....	17,291,000	
	Add flash, $52,300 \times (160 - 118 = 42) \times 0.94 =$	2,065,000	
	Available for evaporation.....	19,356,000	
	L @ 115 = 1027.2. $E_5 = \frac{19,356,000}{1,027.2} =$		18,850
	Transferred to No. 1.....		33,450
	Evaporation No. 1 (assumed).....		17,500
	Concentrate ex. No. 1.....		15,950
(1)	Evaporation by assumption, $17,500 \times 968.4 =$	16,900,000	
	Add heat, $33,450 \times (225 - 118 = 107) \times 0.88 =$	3,150,000	
	Heat requirements S.B. 1.....	20,050,000	
	L @ 240 = 952.1. Steam to No. 1 = $\frac{20,050,000}{952.1} =$...		21,000

HEAT BALANCE, SEXTUPLE, CYCLE (3), SEQUENCE 2-3-4-5-6-1.

Body	Specifications	Heat	Liquor
(2)	Vapor ex. No. 1, 15,100 lbs. @ 218 = $15,100 \times 966.5 =$	14,600,000	100,000
	Ded. heat, $100,000 \times (209 - 175 = 34) \times 0.96 =$	3,260,000	
	Available for evaporation.....	11,340,000	
	L @ 208 = 972.9. $E_2 = \frac{11,340,000}{972.9} =$		11,650
	Transferred to No. 3.....		88,350
(3)	Vapor ex. No. 2.....	11,340,000	
	Add flash, $88,350 \times (209 - 196 = 13) \times 0.955 =$	1,095,000	
	Available for evaporation.....	12,435,000	
	L @ 195 = 980.9. $E_3 = \frac{12,435,000}{980.9} =$		12,700
	Transferred to No. 4.....		75,650
(4)	Vapor ex. No. 3.....	12,435,000	
	Add flash, $75,650 \times (196 - 180 = 16) \times 0.947 =$	1,145,000	
	Available for evaporation.....	13,580,000	
	L @ 179 = 990.5. $E_4 = \frac{13,580,000}{990.5} =$		13,700
	Transferred to No. 5.....		61,950
(5)	Vapor ex. No. 4.....	13,580,000	
	Add flash, $61,950 \times (180 - 157 = 23) \times 0.937 =$	1,337,000	
	Available for evaporation.....	14,917,000	
	L @ 155 = 1004.5. $E_5 = \frac{14,917,000}{1004.5} =$		14,850
	Transferred to No. 6.....		47,100

HEAT BALANCE, SEXTUPLE, CYCLE (3), SEQUENCE 2-3-4-5-6-1
 (Continued)

Body	Specifications	Heat	Liquor
(6)	Vapor ex. No. 5.....	14,917,000	
	Add flash, $47,100 \times (157 - 118 = 34) \times 0.915 =$	1,680,000	
	Available for evaporation.....	16,597,000	
	L @ 115 = 1,027.2. $E6 = \frac{16,597,000}{1,027.2} =$		16,100
	Transferred to No. 1.....		31,000
	Evaporation No. 1 (assumed).....		15,100
	Concentrate ex. No. 1.....		15,900
(1)	Evaporation by assumption, $15,100 \times 966.5 =$	14,600,000	
	Add heat. $31,000 \times (228 - 118 = 110) \times 0.87 =$	2,960,000	
	Heat requirements S.B. 1.....	17,560,000	
	L @ 240 = 952.1. Steam to No. 1 = $\frac{17,560,000}{952.1} =$		18,450

As to the possibility of improving the performance of this cycle over and above that shown in the heat balances, note as follows:

1. Considerable saving of steam can be accomplished by the principle of vapor heating.

With the quadruple effect, the feed entering the second body can be heated with vapors from this unit from 175° to 182° , although the range is so small that it is hardly worthwhile.

With the quintuple effect, vapor heating with No. 2 vapors can be used to advantage, the range being from 175° to 192° .

With the sextuple effect, the liquor can be vapor heated with vapors from the third effect from 175° to 185° , and with vapors from the second from 185° to 198° .

2. In all cycles, vapor heating can be used on the liquor in transit from the last effect to the first, where the range of heating is large, and should therefore be worthwhile.

3. The condensate from the steam belts can be by-passed as before with gain, and no undue complication.

General Discussion. Summarizing this information, below is given a table showing the steam consumption for each set of each cycle, as well as the load on the condenser, which is the same as the evaporation in the last body.

Nomenclature	Steam Consumption			Load on Condenser		
	Cyc. 1	Cyc. 2	Cyc. 3	Cyc. 1	Cyc. 2	Cyc. 3
Finisher	5,670					
Evaporator	23,350					
Total, quadruple	29,020	24,400	25,100	21,900	22,950	22,950
Finisher	5,670					
Evaporator	19,400					
Total, quintuple	25,070	20,100	21,000	18,150	18,550	18,850
Finisher	5,670					
Evaporator	16,900					
Total, sextuple	22,570	17,270	18,450	15,600	15,600	16,100

Cycle 2, the counter current scheme seems to have the better of the argument from the standpoint of steam consumption, which was expected. There is one objection to this cycle in the matter of the inevitable intermediate feed pumps between each body of the multiple effect required to transfer the liquor from a higher to a lower vacuum.

The economy of cycle 3 is not very much lower, and does away with this complication, and must therefore be given consideration. In many of these solutions, pumps give a great deal of trouble and are difficult to keep up. The packing of the stuffing boxes seems to be a source of unceasing work. Especially where such pumps have to draw from a vacuum, the annoyance will be serious.

As regards the load on the condenser, there is very little to choose between cycle 2 and cycle 3.

Cycle 1 uses more steam than either of the other two, and the load on the condenser is greater in every case. It is no doubt simpler in operation, and when improved by the suggestions made, will give good work.

In the establishment of the temperature drops, there was given an estimate of the probable coefficients of heat transmission based upon certain data bearing upon the subject. Such coefficients naturally apply only to the apparatus on which they were established, and an estimate can be made of the amount of heating surface required under these conditions. Whereas, such an estimate would not apply in the average case wherein the type of evaporator would be different, still a good idea can be obtained of the proportional amounts of heating surfaces that would be required for the various cycles. The surfaces must be proportioned to the transmitted heat in each case, and not in the ratio of the evaporation as was assumed in making the preliminary calculations.

The theoretical surface determined in this way must be discounted very materially for the conditions obtaining in practice and which were not allowed for in the previous calculations. These include the tendency to foul, gas binding, lowered rate of evaporation, and the like. It would seem fair to use about twice the surface thus determined, and the log shows this amount. The surfaces established are not the same for all bodies according to the estimate. This is due to the large heating operation in certain bodies according to their cycles. In practice, the surfaces are generally made the same in all bodies. Where there are large heating operations, it is better to do them in heaters designed for the purpose, and this will greatly assist in the successful operation of the evaporator.

The ratings shown are probably safe where the apparatus is kept in a reasonably clean condition by periodic cleanings, depending upon the conditions as encountered at the plant and locality and process.

In considering the steam economy of these various cycles it is necessary to consider also the cost of the apparatus required to accomplish the work. No attempt will be made here to make up estimates of cost which will always be subject to great variations and would therefore be misleading. It is interesting to point out, however, that the heating surfaces required by cycle 1 are very materially lower than those for the other combinations.

EVAPORATION

EVAPORATING SURFACES. CYCLE (1).

Evaporating Set	Bodies	Transmitted Heat (See Balances)	Net Temp. Drop	Heat Transmission	Theoretical Surface	Double for Inefficiency
Finisher		5,390,000	30.0	585	306	612
Quadruple	1	22,200,000	10.5	1,257	1,680	
	2	17,110,000	14.5	1,000	1,180	
	3	18,515,000	22.0	720	1,170	
	4	19,995,000	52.0	330	1,165	
Total					5,195	10,390
Quintuple	1	18,470,000	7.5	1,280	1,920	
	2	13,100,000	10.3	1,030	1,235	
	3	14,267,000	14.2	813	1,240	
	4	15,485,000	19.5	643	1,235	
	5	16,679,000	43.8	307	1,210	
Total					6,840	13,680
Sextuple	1	16,100,000	5.5	1,280	2,290	
	2	10,535,000	7.3	1,070	1,350	
	3	11,385,000	9.5	897	1,335	
	4	12,338,000	12.5	740	1,332	
	5	13,308,000	17.1	578	1,350	
	6	14,387,000	38.3	277	1,355	
Total					9,012	18,024

EVAPORATING SURFACES. CYCLE (2). COUNTER CURRENT.

Evaporating Set	Bodies	Transmitted Heat (See Balances)	Net Temp. Drop	Heat Transmission	Theoretical Surface	Double for Inefficiency
Quadruple	1	23,249,000	24.6	575	1,660	
	2	22,077,000	21.0	623	1,685	
	3	20,525,000	22.4	535	1,710	
	4	17,890,000	25.6	426	1,645	
Total					6,700	13,400
Quintuple	1	19,169,000	18.7	585	1,750	
	2	18,304,000	15.1	674	1,795	
	3	17,250,000	16.7	562	1,835	
	4	15,710,000	18.0	480	1,820	
	5	13,390,000	19.7	397	1,710	
Total					8,910	17,820
Sextuple	1	16,439,000	15.4	591	1,810	
	2	15,762,000	11.9	718	1,845	
	3	15,023,000	12.8	620	1,890	
	4	13,930,000	13.8	527	1,920	
	5	12,460,000	14.9	451	1,860	
	6	10,370,000	15.4	396	1,700	
Total					11,025	22,050

EVAPORATING SURFACES. CYCLE (3).

Evaporating Set	Bodies	Transmitted Heat (See Balances)	Net Temp. Drop	Heat Transmission	Theoretical Surface	Double for Inefficiency
Quadruple	1	23,910,000	18.7	591	2,170	
	2	20,500,000	15.7	776	1,680	
	3	18,773,000	22.1	602	1,410	
	4	20,848,000	37.7	383	1,445	
Total					6,705	13,410
Quintuple	1	20,050,000	13.3	602	2,500	
	2	16,900,000	10.5	835	1,925	
	3	14,210,000	14.1	680	1,485	
	4	15,655,000	19.9	521	1,510	
	5	17,291,000	31.2	358	1,550	
Total					8,970	17,940
Sextuple	1	17,560,000	10.2	613	2,810	
	2	14,600,000	7.6	885	2,170	
	3	11,340,000	9.6	762	1,550	
	4	12,435,000	12.7	630	1,555	
	5	13,580,000	17.3	498	1,580	
	6	14,917,000	26.7	345	1,620	
Total					11,285	22,570

Additional Notes. As to the type of apparatus used, it is probable that better results will be obtained by the so-called long tube design, although where there is no tendency to foam, good results are obtainable with the horizontal tube evaporator, which has the additional advantage that tubes can be removed and replaced easily.

It is well to bear in mind that in many cases there are non-condensable gases given off during evaporation, and that such gases are corrosive. If, therefore, thorough provisions are not made for their continuous removal, there will be tube failures, besides the other ill effects of gas binding on the heating surfaces. Efficient cycles are not possible unless the steam chests of all evaporators are maintained free from these gases.

As has been mentioned in the chapter on Beet Sugar, if there is a considerable quantity of live steam available, and a shortage of exhaust, it is possible to utilize the principle of vapor compression, which can be superposed upon the cycles outlined, preferably, however, on cycle 1, where operation at or above the atmosphere takes place, with low densities, and therefore small temperature differences, permitting high ratios of compression per pound of live steam used. These vapors, however, are possibly of a corrosive nature, and erosion of the throats of the compressors will be serious, and these should therefore be made of acid-resisting material. The same applies to mechanical compressors if these are used instead of thermo-compressors.

The economies possible by the use of this method will depend upon the range of pressures and the conditions of operation imposed.

Chapter 8.

Delicate Materials.

The successful evaporation of very delicate materials requires careful study and precautions. In this group may be mentioned milk, gelatine, and similar materials.

Such solutions are very sensitive to heat, and are adversely affected by excessive temperatures. For this work, the evaporators must be designed so that the length of time during which the material remains in contact with the heating surface is maintained at an absolute minimum. This is accomplished by cutting down the liquor contents of the evaporator to what is absolutely necessary and no more, and using closely grouped heating surfaces. Several types of apparatus are adapted to this work. The Yaryan effect, on account of its extremely low quantity of liquor in transit, was at one time considered standard. Since then, however, several other types have been used with success, and all, or nearly all, have the same general characteristics.

It is well to note that in many cases the cost of steam to do the work becomes a minor consideration, as compared with the quality of the product, and for very delicate solutions, it pays to run single effect with steam at sub-atmospheric pressure, and consequently much lower temperature.

The solution being evaporated in an apparatus is not all the same temperature, as has been pointed out in the chapter on Natural Circulation, and according to design and method of operation, this temperature at some parts may reach a point considerably above the theoretical, sufficiently so as to injure the materials seriously, hence the justification for the use of steam at sub-atmospheric pressure. In the Yaryan effect, this is probably true, though the time element is undoubtedly shorter than in any other design.

As to materials of construction, it is customary to use either aluminum, Monel, nickel, or tin lining. Aluminum provides the cheapest construction, and is satisfactory as long as the solution to be treated is in an acid condition. If, however, there is a possibility of alkalinity, aluminum is unsuited.

Operation by the continuous batch system is quite satisfactory, and permits accurate control of the density.

In the concentration of milk, very special precautions are necessary. In the first place, all apparatus must be thoroughly sterilized before beginning, and cleaned afterwards. In the green state, milk gives off a small amount of gas, which causes foaming when ebullition is started. In addition, it contains albumens which coagulate very readily when heated.

The standard method of concentrating milk is first to preheat it, driving off the gas, after which it is concentrated in coil vacuum pans. In operation, these coils become coated with a white material at every batch, presumably coagulated albumens. This has to be removed by sand papering or some other approved method after each operation. It is very doubtful if this is the best way to accomplish the work. It would seem to be better to use a low liquor content and quick evaporation without preheating, with the continuous batch as suggested above. Injury to the albumens would thus be avoided or minimized, particularly if low pressure steam is used.

This method of operation necessarily involves low ratings of evaporation, on account of the low temperature difference between the boiling liquid on one side, and the low temperature steam on the other. It is a fact, however, that the surfaces do not foul as rapidly, and the speeds though lower, can be maintained through a much longer period of time. In the long run, it is questionable if there is any loss of time, and certainly, the product has been subjected to less injury.

With these low temperatures and high vacua, great care must be used in providing against entrainment losses. Of course, with the long tube type of evaporator, there will be no foam, but the liquid leaves the tubes at high speed, sometimes as high as 200 feet per second, and is in a state of very finely divided drops which it is difficult to stop. The task is far from being impossible, however, and the excellent results fully justify the pains.

One other precaution is to maintain a sufficiently high rate of liquor circulation with such evaporators in order to avoid local overconcentration, which would quickly plug the tubes so affected. It is not always advisable to try to get the maximum evaporative rates by cutting down the liquor circulation, for it will always be found that under such conditions, these rates will not be maintained. The final result is the determining criterion.

Chapter 9.

The Heavy Group.

There is a group of special solutions, such as caustic soda, caustic potash, and calcium chloride, that requires special elaboration. In all of the above, the solutions carry salts, usually chlorides, which have to be removed as the concentration goes on. The usual provisions for salt removal must be made as outlined in the chapter on Crystallizing Solutions.

As evaporation progresses, and salt is crystallized out, the concentration increases, the gravity building up, and the viscosity and boiling point rising, reaching an ultimate final in the evaporator of perhaps 45 to 50 per cent. solids, when further evaporation becomes difficult.

Concentration beyond this point is done in direct fired cast iron pots. It is impossible in the evaporator with steam on account of high temperatures. The range of boiling points from initial to final concentration is very considerable. In the case of caustic soda, the boiling point raising will begin with about plus 15° F. (brine from electrolytic cells), and end with plus 70° F. When operating in double effect, which is customary in this industry, it will be seen that a very considerable part of the available temperature fall is swallowed up in taking care of these boiling temperature losses.

Perhaps the preferable scheme is the so-called reversible double effect. Calling the two bodies A and B, the concentration is carried on in the direction of A to B until the available temperature difference has been so encroached upon as to very materially slow down the apparatus. Then A and B are run as single effects until the desired concentration is reached in B, when it is dumped, new liquor introduced, and operation resumed as a double effect in the direction of B to A, repeating the cycle at the proper time. The only complication involved is one of piping.

If sufficiently high pressures are available, A and B can be run as a straight double effect. In some large plants, there may be six or eight bodies, which can be cut in and out as first effects, second effects, or single effects of a general system, as desired according to the stage of concentration.

High vacua help very materially in the concentration of these high boiling solutions, as in this way the available temperature drop is greatly increased. The possibility of high vacua is entirely dependent upon the temperature and quantity of injection water for the condensers, which is usually a seasonal factor, especially in southern climates.

Materials of construction are cast iron, wrought iron, nickel and char-

coal iron tubes. Caustic has a very curious action on wrought iron used in the tubes. After a certain period, the metal becomes very brittle, and fails by cracking, although when originally installed, it might have been perfectly ductile. It is very advisable to select the best of materials.

The joints in the various parts are very difficult to keep tight and must be made with material that will withstand the action of the solution. Stuffing boxes are a source of continuous trouble and should be avoided wherever possible.

It has been customary to use the basket type evaporator for this work, and to remove the salt by means of traps, where it can be washed free of contamination before removal.

The ratings are very low on the average, probably from 4 to 5 pounds per square foot per hour.

Chapter 10.

The Evaporation of Glycerine Liquors* and the Wurster & Sanger Evaporator.

Contributed by OSCAR H. WURSTER,
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Sources of Glycerine. All glycerine is recovered from dilute solutions. Practically the entire world's production of glycerine at present is obtained from two general classes of glycerine solutions, spent soap lye and sweet water. Spent soap lye is obtained upon the saponification of fats or oils by means of caustic soda in the manufacture of soap. Sweet water is obtained by the direct splitting of fats and oils into glycerine and fatty acids, usually by the autoclave process or the Twitchell process. These processes are used in the production of fatty acids to be used for the manufacture of stearic acid and red oil, of distilled fatty acids, and of soap.

In soap kettle practice fats and oils are saponified by boiling with caustic soda solution and there are formed glycerine and a sodium soap. After the glyceride is thoroughly saponified, sodium chloride is added to the charge in the kettle in order to separate the soap. Sodium soap is soluble in water, but is practically insoluble in brine. The charge then separates into three layers: an upper layer of soap, an intermediate layer of nigre made up of soap, water and impurities, and a bottom layer of spent soap lye which contains most of the glycerine.

Evaporation of Glycerine Liquors. Spent soap lyes as they are drawn from the soap kettle consist, in general, of glycerine, sodium chloride, caustic soda, sodium carbonate, soap, fatty acids and water, in addition to albuminous and oleaginous matter. The proportions of these ingredients vary considerably with kettle practice and also with the various stocks saponified. The most valuable ingredient is glycerine, the percentage of which will vary with the stock used and the amount of lye drawn from the kettle. Practically all spent soap lyes which are concentrated for glycerine will contain from 2 per cent. to 9 per cent. of glycerol, the average being around 5 per cent.

* Author's Note.—This excellent chapter was contributed by Wurster & Sanger for the fourth section of this book, but was transferred by the author to the third section where it properly belongs. The reader will understand that the description of the apparatus and its claims for special merit are those of the designer, and that at times the points brought out may be at variance with the balance of the text. It has been retained practically intact, however, as in the case of other contributions.

Sweet water from the fat splitting processes will usually contain from 10 per cent. to 15 per cent. glycerol, small quantities of fatty acids and organic matter from the fats and mineral matter from the catalysers and reagents used.

The glycerol content of these liquors is to some extent under the control of the manufacturer. The price of glycerine determines the extent and thoroughness of recovery which is desirable. In either case both the cost of carrying further the recovery of the glycerine in dilute form as well as the subsequent evaporation of such more dilute solutions becomes more costly as the total yield of glycerine is increased. Under any given set of conditions, it is necessary to determine to what extent the recovery can be carried economically and profitably.

As both soap lyes and sweet waters contain, in addition to glycerine, other ingredients which to the glycerine refiner constitute impurities, it becomes necessary to treat these liquors before concentration. Soap lye is to be concentrated to standard 80 per cent. Soap Lye Crude Glycerine, which usually contains 80 per cent. to 84 per cent. glycerol, about 10 per cent. ash and about 3 per cent. organic residue. Sweet water is concentrated to 88 per cent. Saponification Crude Glycerine which usually contains less than 0.5 per cent. ash. Unless the liquors are properly treated and filtered, these standards will not be reached by the finished crude glycerine. An untreated soap lye, on concentration, will yield a crude glycerine analyzing about 40 per cent. glycerol.*

Bearing of Composition of Liquors on Evaporator Design. From the standpoint of the subsequent evaporation, and considering the liquors after the removal of impurities by precipitation and filtration, the essential characteristics of these two glycerine liquors, as well as their differences, are as follows; spent soap lye contains from 10 per cent. to 15 per cent. salt, most of which is precipitated during evaporation, and sweet-water contains scale forming mineral salts in small amounts. These characteristics are reflected in the finished crude glycerine, for, as stated above, standard soap lye crude contains 80 per cent. glycerol, 3 per cent. organic residue and 10 per cent. ash (it is a saturated salt solution), whereas saponification crude contains 0.3 per cent. to 0.5 per cent. ash and 88 per cent. glycerol.

The above characteristics of the liquors to be evaporated have an important bearing on the design of the evaporators. For spent soap lye evaporation provision must be made for the removal of the salt which precipitates during concentration, preferably without interruption of the evaporation. In the case of sweet water, there is no removal of highly soluble salts, as sodium chloride, during evaporation, but there is likely to be some precipitation on the tubes of slightly soluble scale forming calcium salts, and therefore the evaporator design should be such as to permit easy access to the tubes for cleaning.

Early Methods of Evaporation. As in some other industries, evaporation of glycerine liquors was first carried out in steam heated open

*For details of treatment of glycerine liquors see Sanger, *Chem. & Met. Eng.*, Vol. 26, No. 26, June 28, 1922.

pans under atmospheric pressure. There were, of course, numerous objections to this procedure. More equipment and more floor space were required to evaporate a given quantity of liquors as the rate of heat transfer, and consequently the rate of evaporation, was relatively low. The large volumes of steam produced were objectionable in the plant and such pans were often placed outdoors or in special sheds. It was a difficult and slow procedure to reach the higher concentrations and high pressure steam was required. At the high temperatures required, there was a greater loss of glycerine than in vacuum evaporation. Due to lack of circulation of the

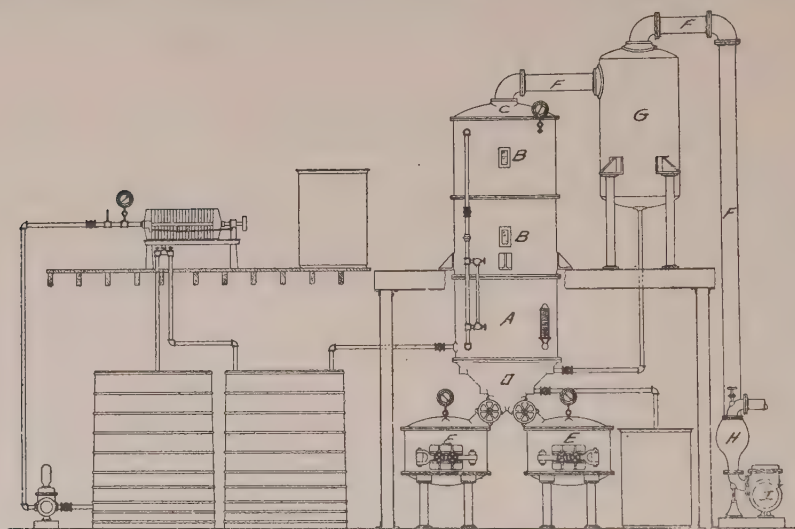


FIG. 1.—Wurster & Sanger Soap Lye Crude Glycerine Plant.

liquid, salt deposited on the heating tubes, thus insulating them and reducing their capacity. The tubes were not easily accessible and were difficult to clean. There was no simple and convenient method of removing and washing the precipitated salt in the evaporation of soap lyes.

Vacuum Evaporation. Early in the history of glycerine recovery soap manufacturers adopted the standard vertical tube type of evaporator operating under vacuum, and this is the most common type now used. Many improvements have been made in the design of glycerine evaporators, but all of them now operate on the basic principles of vacuum evaporation.

The evaporation of glycerine solutions is confined to the use of single and double effect evaporators. The single effect evaporator is lower in initial cost, but requires approximately twice as much steam and condensing water as a double-effect doing the same work.

Fig. 1 shows a simple arrangement of a small spent soap lye treating plant and Wurster & Sanger soap lye evaporator. Fig. 2 shows a similar arrangement for the treatment and evaporation of glycerine sweet water.

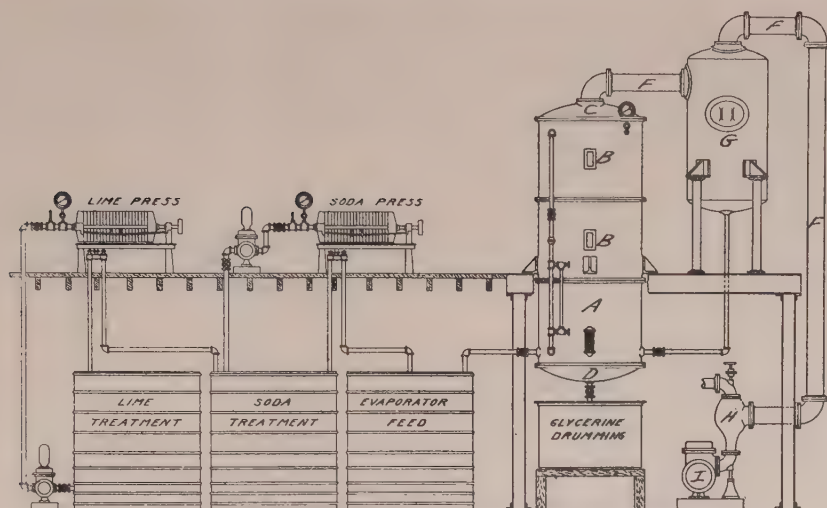


FIG. 2.—Wurster & Sanger Saponification Crude Glycerine Plant.

Description of Wurster & Sanger Evaporator. The evaporator unit consists of the calandria or steam chest *A*, the vapor belt consisting of two rings *BB*, the cover *C*, and bottom *D*. It will be noted that the

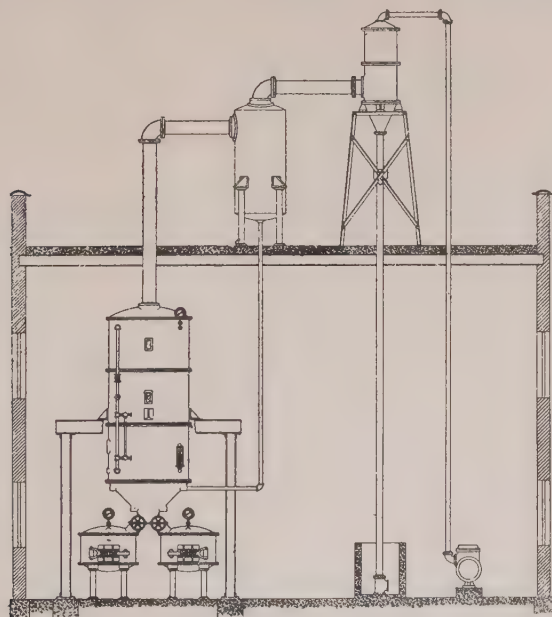


FIG. 3.—Wurster & Sanger Single Effect Crystallizing Evaporator.

soap lye evaporator (Fig. 1) has a conical bottom to facilitate the removal of the precipitated salt, whereas the sweet water evaporator (Fig. 2) has a dished bottom. The soap lye evaporator is equipped with two salt extractors *EE*. The balance of the equipment is identical and consists of the vapor pipe *FFF*, the catchall *G*, the condenser *H*, and the vacuum pump *I*.

In both these installations the catchall is on a level with the evaporator proper, and a jet condenser is used in conjunction with a wet vacuum pump. Fig. 3 shows a crystallizing evaporator with the catchall above the evaporator, a barometric condenser and dry vacuum pump.

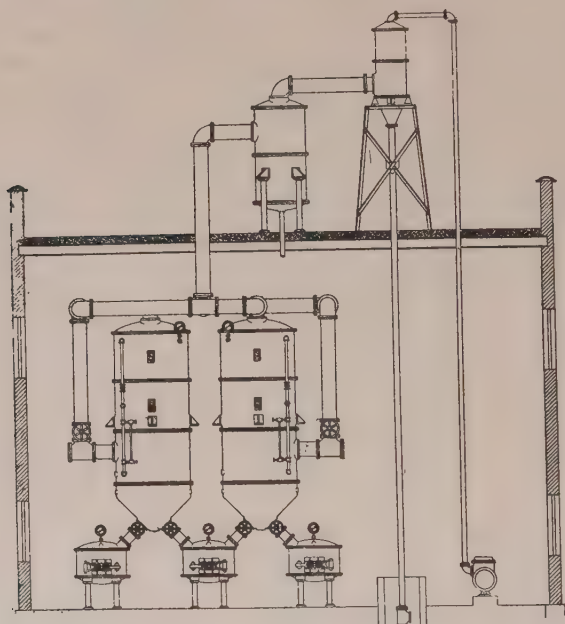


FIG. 4.—Reversible Double Effect Crystallizing Evaporator Unit.

Fig. 4 shows a Wurster & Sanger Double Effect Evaporator, together with the appurtenant equipment outlined above.

All of these evaporators are provided with the necessary gauge glasses, sight glasses, vacuum and pressure gauges, thermometers, liquor lines and other connections.

The general design and arrangement of the Wurster & Sanger Evaporator* is shown in Fig. 5. The calandria or steam chest consists of a cast iron cylinder with an arcuate partition near one side. Copper tube sheets are bolted to the faced flanges of the cast iron ring and copper tubes inserted. The general construction of this calandria is clearly shown in

* Sanger, U. S. Patent 1,508,130, Sept. 9, 1924.

Fig. 6, giving two views of this important part of the evaporator. The vapor belt is provided with baffles which deflect downwards the liquor which is carried up with the steam.

Advantages of the Design of the Wurster & Sanger Evaporator. The design of this evaporator offers a number of natural advantages over the usual forms of evaporators. It also permits of improved construction

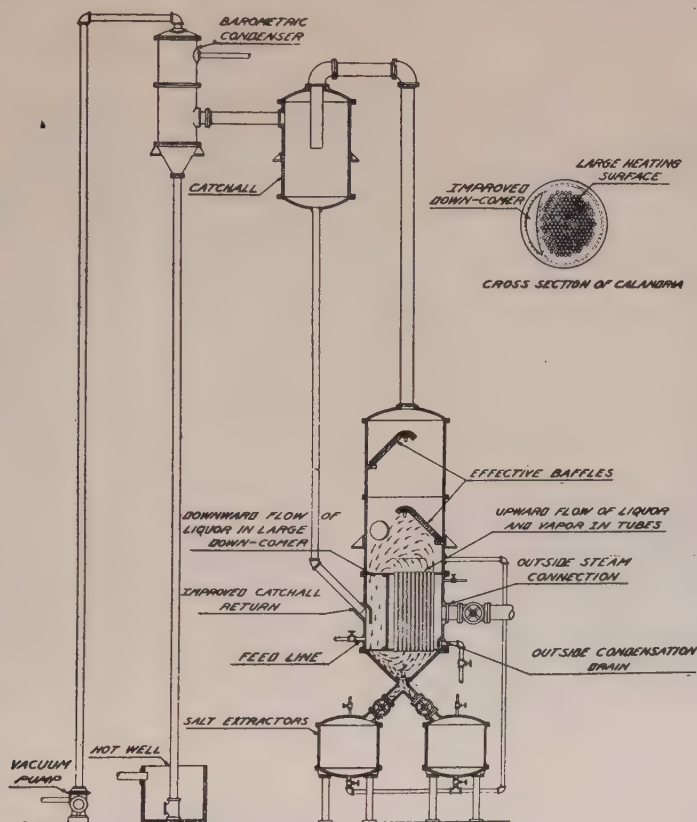


FIG. 5.—Wurster & Sanger Evaporator.

and arrangement in several respects. In the calandria the heating tubes are all nested on one side of the curved dividing partition. The liquor, on boiling and evaporating, therefore rises on that side of the evaporator. The open space on the other side of the partition, the down-comer, is a large unrestricted area. The arcuate wall in common with the steam chest is farthest from the steam inlet, and is therefore relatively cool. The other wall is the outer unheated wall of the evaporator. The liquid carried up in the tubes flows down freely and without obstruction or resistance in this large down-take. There is thus set up a natural circular movement

of the liquid within the evaporator, upwards through the heating tubes and down the down-comer chamber. This free, positive, rapid circulation in a definite direction is in marked contrast to the general churning effect produced in the older forms of evaporators. The advantages of this naturally directed flow of liquor within the evaporator are several. Longer tubes may be used without sacrificing efficiency. The circulation is so effective that the counter tendency of the liquid to drop back in the tubes as the gravity increases is off-set. The rapid circulation conduces to a higher rate of heat transfer and hence a higher rate of evaporation. The



FIG. 6.—Calandrias for Wurster & Sanger Evaporators.

elimination of the churning effect so general in other evaporators results in a reduced grinding effect of the salt particles on each other, with the result that the crystals grow more freely. The larger crystals settle out in the cone bottom and pass on to the salt extractors. In the salt extractors the coarsely crystallized salt is more thoroughly and rapidly freed of glycerine than the fine crystals formerly produced.

Disadvantages of Some of the Older Designs. The advantages of the large unrestricted down-comer on one side and free from friction and upward currents, over a large center tube down-comer are those enumerated above which result in giving the liquid definite direction and a high velocity over the heating surface. With this improved down-comer the direction of flow is assisted by a proper arrangement of baffles as shown

(Fig. 5). The liquid cannot be so simply and positively direct to and down through a centrally located down-comer, and at the same time permit a free escape of the vapors. In other words, the large center down-comer arrangement gives rise to the lack of direction and the general churning effect referred to above. The center down-comer also has its entire surface exposed to the heating chamber and therefore upward currents are produced within the down-comer, thereby reducing its efficiency.

The basket type of calandria offers as a down-comer an annular ring between the basket and the outer shell. This space appears of adequate area, but brackets necessary to support the basket offer some obstruction. The more serious objections include the general fault cited above of lack of specific direction of flow of the liquor with the resultant churning effect and its accompanying difficulties already explained. Other difficulties and objections are largely mechanical, but none the less serious in the operation and maintenance of the equipment. The only way to reach the basket steam chest is through the outer wall of the calandria. There is, therefore, a connecting steam pipe from the basket to an opening in the outer shell. Consequently there are joints to be packed and kept tight within the evaporator. The necessary expansion and contraction with heating and cooling of this connecting piping is a severe strain on the joints and basket, making it difficult to avoid leaks and possible losses. The condensation drain from the basket must also pass through the outer shell and involves internal piping and joints. With slight but continuous leaks in these internal connections, the total loss of glycerine may easily become serious. The repairing and replacing of the basket type calandria is more difficult and expensive than in the case of the improved form in which all connections are made on the outside.

Further Advantages of the Wurster & Sanger Evaporator. The design of the Wurster & Sanger calandria and evaporator is such that no baffling of the steam in the calandria is necessary to produce the best results. The steam enters at the side farthest from the down-comer. There is therefore an automatically decreasing rate of evaporation from the point of steam inlet to the partition wall of the down-comer chamber. Thus the most efficient use of steam and the most effective circulation of the steam and the liquor in the evaporator are brought about naturally by the design.

To obtain the best results in evaporation it is an advantage to have the passage of the gases, vapors and liquids over the heating surface as rapid as possible. This end is accomplished by the natural and positive circulation which results from the design of our evaporator.

This free and unimpaired circulation is also a marked advantage in keeping the tubes clean and working the separated salt at once to the bottom of the evaporator.

Other improvements of this design of calandria referred to above result from the fact that direct connections can be made on the outside shell of the calandria section to both the steam chest and to the liquor section. This is the only design so far developed to make this possible. The steam connection is made on the outside and opens directly into the

steam chest. There is no inside steam piping to line up, no steam connection from an outer shell to an inner basket. The condensation drain from the calandria is also made directly to the outer shell. The catchall return can be connected direct to the down-comer chamber where it will be sealed by the liquor in the evaporator against vapor passing up the return line, and will yet be high enough that it will not become clogged with salt.

Copper and Steel Tubes and Tube Sheets. There arises occasionally a discussion of the relative merits of copper and steel tubes and tube sheets. Copper tubes are more efficient in the transfer of heat, have a longer life and give better service under normal operating conditions in the evaporation of glycerine liquors. The conditions for an electrolytic action between the copper and cast iron are apparently present, but no deleterious effect from such action has come to our attention. There have been occasional instances of a rapid development of leaking tubes. In no case in our experience has this been traced to electrolytic action. It has usually been difficult to positively determine the cause, the trouble usually disappearing as unexplainably as it appeared. A probable cause has been the occasional occurrence of sulphides in the caustic used in saponifying the stock.

Other Parts of the Complete Unit. The catchall is used for decreasing losses by entrainment and also as a separator if a charge is pulled over from the evaporator. The large vapor space over the calandria in the Wurster & Sanger evaporator is in itself a safeguard against such losses. The catchall is equipped with a return drain line to the evaporator and also with a water-gage column. There appears in practice to be some advantage in placing the catchall at an elevation above the evaporator as shown in Figures 3, 4 and 5.

The condenser may be of the jet type directly connected to the vacuum pump as shown in Figs. 1 and 2, or of the barometric type as shown in Figs. 3, 4 and 5. The barometric condenser serves to maintain a more even vacuum than is obtained with the jet type, requires a smaller vacuum pump than when the condensing water is passed through the pump and is used on practically all large installations.

The salt extractors shown in Figs. 1, 3, 4 and 5 are all designed to steam out the salt and return the glycerine liquor direct to the evaporator through a connecting drain line as shown in Fig. 5. With careful steaming and draining, and a good crystallization of the salt, the glycerine content of the salt will be kept down to about 2 per cent. Frequently, under unfavorable conditions, the salt will show on analysis a glycerine content up to 4 or even 6 per cent. It is not unusual to see some glycerine oozing from recovered salt in a storage bin or on the kettle room floor. It is often argued that the glycerine returned to the kettles in the salt is again recovered. It can be shown, however, that much of the glycerine so returned is permanently lost and the loss is appreciable even with only 2 per cent. glycerine in the salt.*

* Of the glycerine that is returned to the kettle with salt, using recovered salt in all the changes, approximately 40 per cent. remains in the soap. This is so because from the last changes only a relatively low recovery is made, these changes having

More recent practice in large plants is to dispense with this old form of salt extractor and substitute a glycerine drum. The salt sludge is allowed to collect in this drum until it is full, and then the sludge is blown to a centrifugal separator where the glycerine is completely removed from the salt. This procedure may be considered as representing the present best practice in the industry for the handling and recovery of the precipitated salt.

Operation of Evaporators. The general principles of single and double effect evaporators, their arrangement, and methods of operation being understood by the reader, minute operating instructions for the evaporation of glycerine liquors will not be given here.* The general operation of the Wurster & Sanger evaporators for the evaporation of glycerine liquors may be briefly described.

The procedure in single effect evaporation of spent soap lyes is as follows: Vacuum is pulled on the equipment and the treated and filtered lye is fed to the evaporator. Steam and condensing water are turned on and as evaporation proceeds more lye is fed to the evaporator. As the salt crystallizes out, it is removed through salt extractors or salt drums. Low pressure steam up to 5 lbs. is adequate for the concentration to semi-crude. For concentration from semi-crude to 80 per cent. crude, steam up to about 25 lbs. pressure is usually used.

In the concentration of glycerine sweet water the same general procedure is followed. There is no salt to remove. The final concentration to 88 per cent. Saponification Crude is also made with a steam pressure on the calandria of about 25 lbs.

In double-effect evaporation vacuum is pulled on the entire equipment. Both effects are then charged with liquor. Steam is turned on the first effect and condensing water turned on the condenser. After the first effect is brought to boiling, the vapor from this effect passes on to the calandria of the second effect and the liquor in this effect in turn starts to boil. The second effect is usually operated under 25 to 28 inches vacuum and the first effect about 10 inches lower. As the liquor boils off, the second effect is fed from the first and the first effect from the feed tank to maintain the desired liquor levels in each case. When the con-

fewer washes after the glycerine has been added. If any recovered salt is used in graining soap from which the lyes are not to be used the loss is, of course, greater.

The centrifugal in ordinary practice with an exceedingly small volume of wash water brings the glycerine in the salt below 0.25 per cent., whereas the usual form of salt box or vacuum filter requires considerable wash to bring it down to 1 per cent., and in practice it is seldom kept down so low. It usually runs from 2 to 4 per cent. If the difference in glycerine content between the centrifuged salt and the filtered salt is figured as 2 per cent, and assuming 40 per cent. of this remaining in the soap, we have 0.8 per cent. glycerine on the salt used. Further assuming 12 per cent. salt in average lye and 3 lbs. of lye per pound of fat, or 36 per cent. salt on the fat, there is a saving of 0.288 per cent. glycerine based on fat due to centrifugal drying. This equals 0.36 per cent. crude glycerine or 3600 pounds crude glycerine from one million pounds of fat.

* See Walter E. Sanger, Recovery of Glycerine from Spent Soap Lyes, *Chem. & Met. Eng.*, Vol. 26, No. 26, June 28, 1922, and Solving Evaporator Problems of the Soap Industry, *Chem. & Met. Eng.*, Vol. 29, No. 11, Sept. 10, 1923.

centration in the second effect reaches the semi-crude stage, the first effect is shut off and the semi-crude in the second effect is finished to crude with live steam. In larger plants the double effect evaporators are usually operated continuously as double effects and a separate and smaller single-effect is used to finish the semi-crude to crude. This arrangement effects a greater economy and efficiency in the use and operation of the equipment.

Glycerine Losses, Causes and Remedies. In the recovery and refining of glycerine, there are many possible sources of loss from the saponification of the fats and oils to the refined dynamite or C. P. glycerine. We are concerned here only with the possible losses which may occur in the evaporation process proper. These may be enumerated as follows:

1. Loss by entrainment.
2. Loss by boiling over.
3. Loss by distillation.
4. Loss by vaporization.

Entrainment losses of glycerine usually consist in the passing over of small particles of the solution being evaporated. These extremely fine drops may pass from the surface of the boiling liquor as such or may be the result of breaking up of larger drops by impingement or of the breaking of bubbles. Comparatively large drops of liquor may be carried away with the vapors from the glycerine evaporator. To reduce such possible entrainment losses to a minimum a large, high vapor chamber with adequate baffles is provided above the calandria and a large catchall is placed in the vapor line ahead of the condenser. There are, therefore, provided two safeguards for returning entrained liquors to the liquor in the evaporator.

A sudden increase in the vacuum with a consequent lowering of the boiling point of the liquor in the evaporator may cause such a sudden generation of steam in the body of the liquor as to lift the entire charge out of the evaporator. The charge may be caught in the catchall or may partly or entirely pass out through the condenser. In the plants having proper equipment, this is the most prolific source of glycerine loss which has come under our observation. This condition is most likely to occur when starting the evaporator in operation and when there is a sudden temporary interruption of the water supply to the condenser. Both these causes can be readily eliminated by careful operation of the equipment.

There is a tendency of glycerine to distill off with the water vapors at temperatures in excess of 170° F. Crude glycerine should be finished under temperature conditions as low as possible.

There is also the possibility of some glycerine loss in the form of glycerine vapor due to the superheating of the liquor in the evaporator. Such superheating has been found to vary from 10° to 32° F. This, however, is a minor source of trouble and is guarded against by the large vapor chamber and the catchall. The evaporation should be carried on uniformly and increases in the steam supply to the calandria should be made gradually.

Glycerine liquors in proper condition and properly treated offer no great difficulties in subsequent handling and concentration in properly designed and constructed equipment. When liquors are old and fermentation has set in, special precautions must be observed or there will be serious losses.

SECTION IV.

TYPES OF EVAPORATORS.

Chapter I.

Introductory.

In the previous sections of this book the general phases of Evaporation have been discussed, as well as the technique of the work involved. In this, the last section, is given a description of the various types of evaporators found in everyday commercial practice. It is manifestly impossible to cover all designs, and the subject matter will therefore be limited to apparatus in general use.

Various manufacturers have been invited to assist in this work, and where such manufacturers agreed to cooperate, they were assigned a certain space to describe one type only, on which they specialize, the understanding being that this does not represent their only design. Acknowledgments are made under each heading, the data received having been retained practically without modification.

In mapping out this work, it has been the author's object to show favoritism to no one, and the selection of manufacturers was made by writing to all those advertising in the Chemical Catalog, assigning one topic to each one.

Whereas at times, the information brought out may be slightly at variance with the text, it has been retained intact, so as to give the reader various points of view. It is most gratifying indeed to note the broad minded spirit of cooperation displayed, without which it would have been impossible to write this book.

Chapter 2.

The Standard Effect.

(Courtesy Joubert & Goslin Machine & Foundry Co.)

We Americans are great on standardization. It is the true basis of our commercial success. We first thoroughly investigate, establish facts which we consider sound, and then standardize on these facts, settled, presumably, once and for all. If our work has been well done, we thus avoid great loss of time in useless discussions, thereby conserving our available energies for the study and solution of more abstruse problems.

Once the standards have been established, it is very difficult to change them, and intentionally so. It is not impossible, however, and when new facts make changes imperative, we change, but we are slow in doing so. Our standards are set by what we find to be simple, practical, and free from trouble, and are usually the result of a thorough sifting operation which may extend over a long period of time.

And so, we have in Evaporation, the Standard Effect, which we consider simple, efficient and practical for all ordinary purposes. For extraordinary purposes, we have special designs, but they represent a small part of the grand total.

The Standard Evaporator is a vertical tube apparatus with relatively short tubes, from 4 to 6 feet long, which offer ruggedness, simplicity, and low first cost.

The cylindrical steam belt is the full diameter of the evaporator, and the full length of the tubes less the thickness of the two tube sheets, which extend the full diameter of the flanges, and are secured between the lower steam belt flange and the flange of the bottom below, and between the upper steam belt flange and the lower flange of the vapor belt above. Sometimes, the joint between the steam belt flange and the tube sheets is made independently by the use of countersunk bolts straddling the main flange bolts, which gives a much better construction, and enables shipment of the assembled steam belt from the shops without danger of disrupting the joint while the steam belt is in transit from the works to the field.

Usually, there is provided in the centre a downtake of rather large proportions, perhaps much too big for the work intended, which is certainly not a serious defect. Some manufacturers use smaller distributed downtakes instead of a large central one, thus avoiding a riveted joint, such downtakes being small enough to be expanded.

Steam is admitted into the steam belt from the side into an opening located centrally between the two flanges at some convenient point. Usually, there is an annular space to receive this steam, and slotted holes between

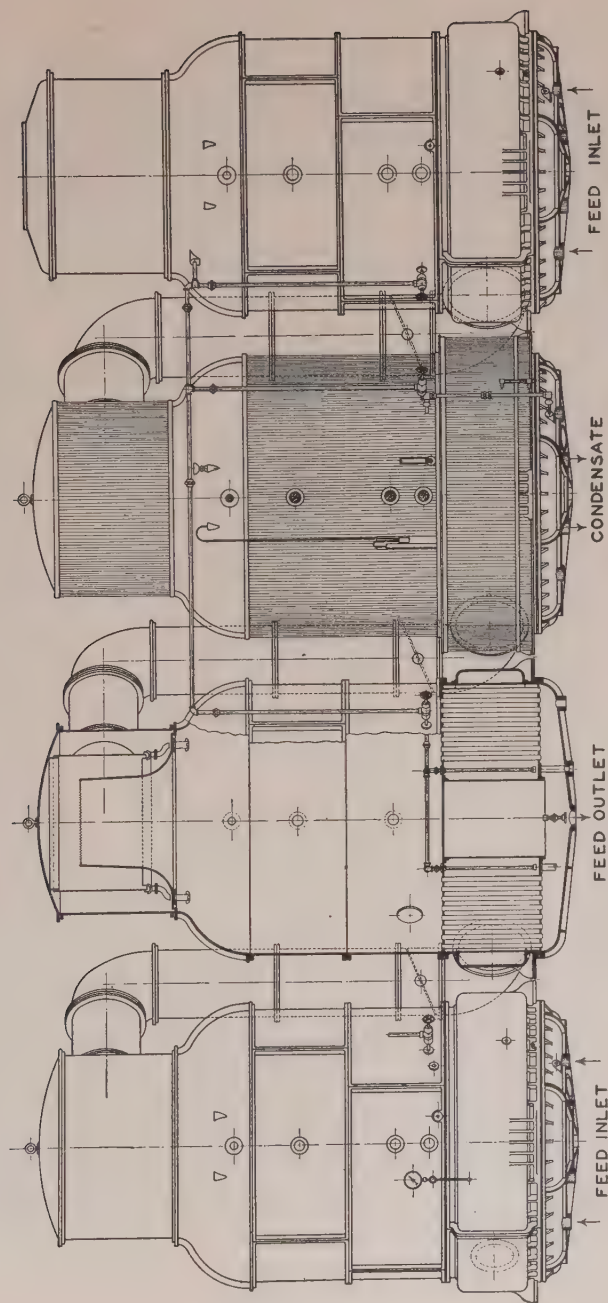
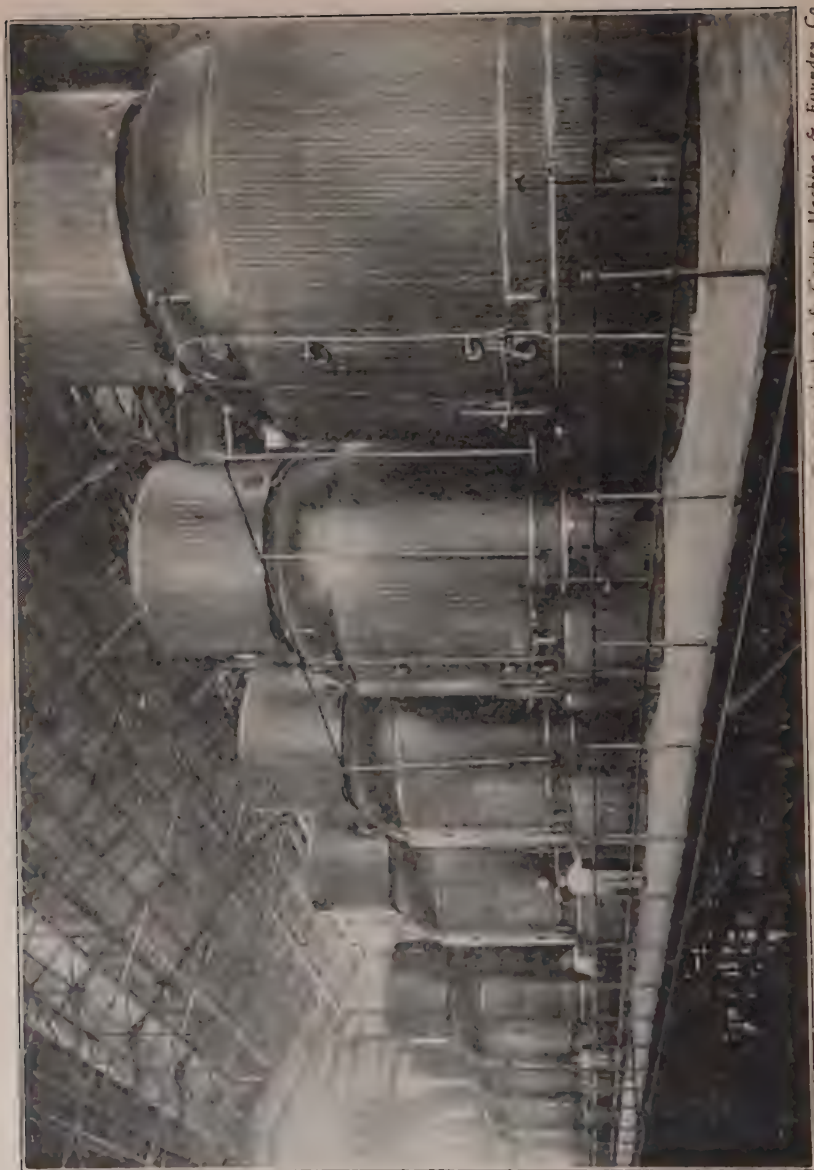


FIG. 1.—Typical Standard Tube Evaporator Showing Bodies Stripped, in Section, and Completely Mounted.
(Joubert & Goslin Machine & Foundry Co.)



Courtesy, Joubert & Godin Machine & Foundry Co.

FIG. 2.

this and the heating surface for proper distribution. (See chapter on Types and Steam Belts.)

The bottom below the steam belt is deep enough for a man to work on the lower tube sheet in expanding tubes or other necessary repair work, and there is a manhole provided for access from the outside. The bottom is provided with openings for feed inlet, feed outlet, condensate outlet, and washout connection.

The vapor space above the steam belt is 8 to 10 feet high, and the full diameter of the body. There should be provided a full set of sight glasses of large diameters, a manhead, thermometer, level gauges, and accessories. On account of prevailing deficiencies of ordinary catchalls, many prefer to have the vapor belts even higher than specified in order to minimize the danger of entrainment losses. (See chapter on Catchalls and Separation.)

The dome or top casting above the vapor belts is 3 to 5 feet high, according to the size of the apparatus, and is in reality only a reducing piece to bring the diameter down from the vapor belt size to the catchall size. It is provided with a vacuum breaker and sight glasses front and back.

The catchall is attached to the dome, and in general practise is of the bonnet type, with outlet to one side, having vapor piping leading to the next effect of the series, or to the condenser in the case of the last body.

The feeding system is for parallel operation, and flows from bottom to bottom (see chapter on Feeding Systems), although automatic feeding systems are being used more and more, as they compensate for inefficiency of the operators.

In most cases, the condensate is removed from each steam belt by a pump, a "sweet water pump" as it is called. Sometimes, the evaporator is high enough to permit this water to flow by gravity to a tank below, through a barometric seal. Still other designers by-pass the condensate from one steam belt to the next through "flash pots," allowing the steam flashed to enter the succeeding steam belt, and the water to go on from one flash pot to the next, removing it all from the last "pot" with only one pump. This method has been fully elaborated in the text.

The venting or non-condensable gas removal is done by running the gas connection into a header leading to the vapor space of the last effect, or to the condenser. This also has been fully elaborated in the chapter on Non-Condensable Gas Removal.

There is a dumping and washout header below the apparatus with connections to each body.

It is customary to provide covering for protection against radiation losses, either hair felt, asbestos, or magnesia, on the outside of which is placed a neat covering of hard wood for mechanical protection, properly secured with metal bands with draw lugs.

The illustrations accompanying show (Fig. 1) a general drawing of a quadruple effect, and (Fig. 2) an installation photograph of the same.

This design or slight modifications of it represents a very large majority of the types of installations usually encountered in industry.

Chapter 3.

The Baffled Evaporator.

(Courtesy U. S. C. I. Pipe & Foundry Co.)

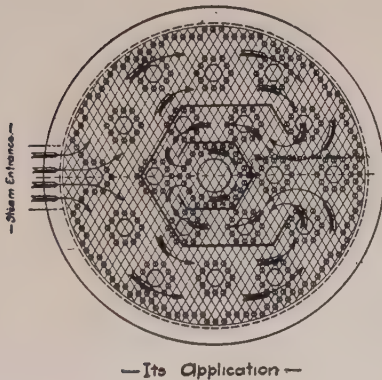
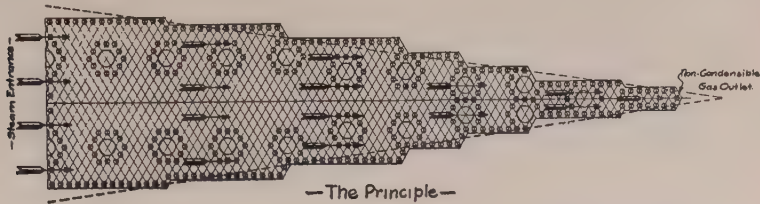
The standard effect with annular steam belt, as described in other pages, represents a design originated perhaps forty years ago. The first vertical tube effects had no annular space, and always gave considerable difficulty in securing satisfactory steam distribution. The idea of the annular space was to provide a reservoir all around the calandria, from which steam would be admitted through slots or perforations, entering in a radial direction, and flowing always towards the centre. Undoubtedly, the scheme was better than the old one, as shown by its almost universal adoption shortly after its introduction.

It is easily conceivable that under fixed conditions, fair results are obtainable, especially in small sized apparatus. Usually, in designing a multiple effect, the calandrias for all the bodies are made alike, as a manufacturing proposition. It is therefore evident that inasmuch as under operating conditions, the volume of steam or vapor admitted in the various bodies will vary with the different vacua or pressures existing, the proportions of both the annular space, and the slotted holes or perforations will not fit all conditions. If the sizes are made to fit last body operation, they will be too large for the others, and if they are designed for first body conditions, they will be too small for the others, and a compromise of proportions will not exactly fit any.

Unless there is an appreciable difference of pressure between the annular space and the calandria, there will be nothing to compel even flow in all orifices. Such a difference of pressure necessarily involves a loss of temperature fall, and a consequent impairment of capacity. In actual design, a compromise is made which presumably involves a minimum of objections, an attempt being made to provide for the flow under last body conditions without excessive friction loss.

It should be noted that especially for large sizes, it is very difficult to provide sufficient cross section in the annular space, and usually, it is found that the cross section of the vapor pipe is considerably in excess of the cross section of the annular space through which the same vapor must flow. In addition, at the entrance to this space, a sudden right angle turn must be made, which always involves pressure losses, especially at high velocities. As a specific illustration, might be mentioned a quadruple effect having a 48-inch vapor pipe between the last two bodies, and an annular space 8 inches by 42 inches high, which represent standard propor-

tions for many designs. The area of the vapor pipe is 1809 square inches, and that of the annular space 332 square inches. The flow of steam through this space being on both sides of the steam inlet, this area must be doubled, making it 664 square inches, which is roughly one third of the vapor pipe area. If twin vapor pipes are used having the same area as the single one in this case, vapor flows into the annular space on both sides of such inlet, and the annular space area must be quadrupled, or 1338 square inches, which is also smaller than the sum of the two vapor pipes.



Typical
—Layout of Tube Sheets—
—The Webre Evaporator—

FIG. 1.

These factors involve high peripheral velocities around the steam belt, which cause a shunting of the slotted holes, most of the flow taking place where the two currents impinge, and there is a local increase of pressure due to the conversion of the velocity head into a static head.

It is evident that if the annular space is designed to meet last body conditions, when it is called upon to meet the conditions existing in other bodies, the distribution of steam flow through the slots will be worse instead of better, for such pressure differential as may have existed will be very much reduced, and most of the steam will enter the calandria at a point opposite the steam inlets. This can actually be observed in large evaporators in operation by lowering the level slightly below normal.

As a result of study and observation of these conditions, many en-

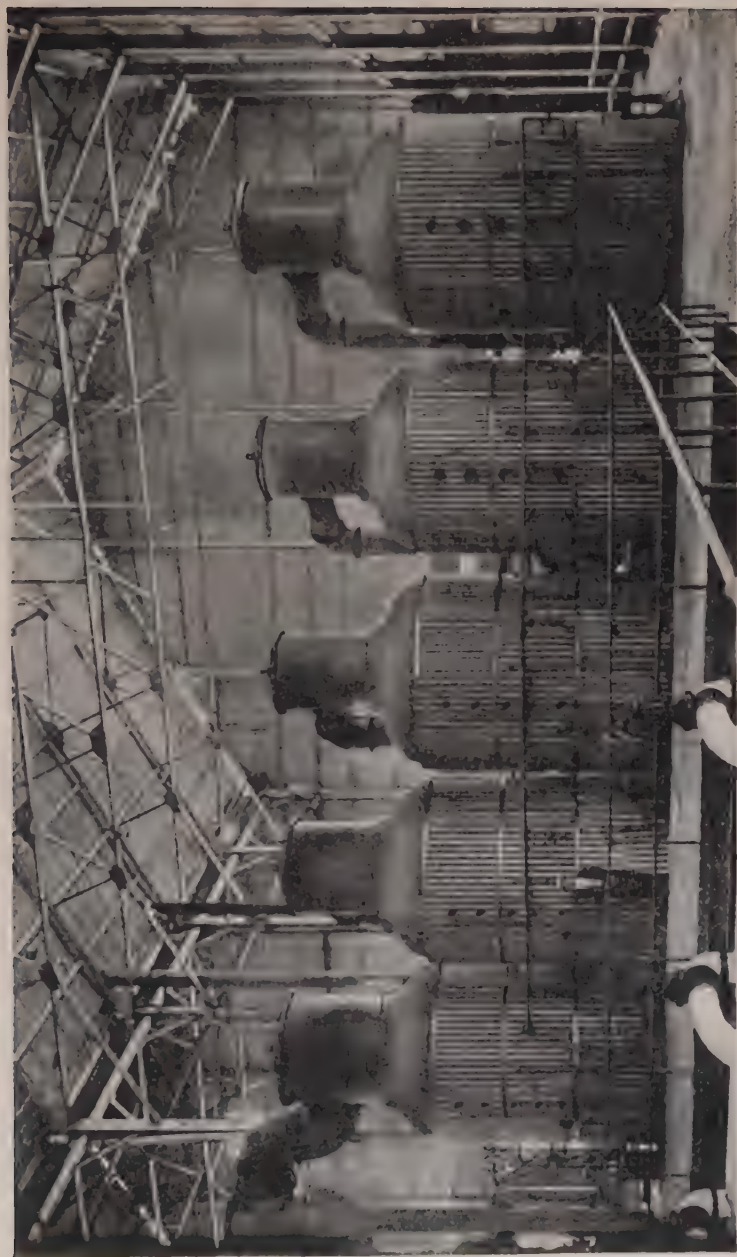


FIG. 3.—Quintuple Effect Webre Baffled Evaporator.

steam is swept to them due to the natural vapor current, and therefore permits a minimum accumulation.

It has been found that evaporators of this design exhibit certain very valuable characteristics. The surfaces are uniformly active throughout, and operate with a very high coefficient of heat transmission. This permits the use of more bodies in series with better economy. It also permits operation at higher ratings than have been heretofore possible, and operation of vapor cells with very small differences of temperature, permitting more economical cycles now becoming more and more necessary with the ever increasing cost of fuel.

Chapter 4.

The Horizontal Tube Evaporator.

Rillieux's first evaporator was built with horizontal tubes, and so, it can be said that the original multiple effect was a horizontal tube apparatus.

Naturally, it has seen a great many developments since its original inception, and many different combinations have been tried. The original apparatus consisted of a horizontal cylindrical shell with horizontal tubes, these being much larger in diameter than the prevailing present practice, and of shorter lengths, of the order of 2 inch diameter, by 12 feet long.

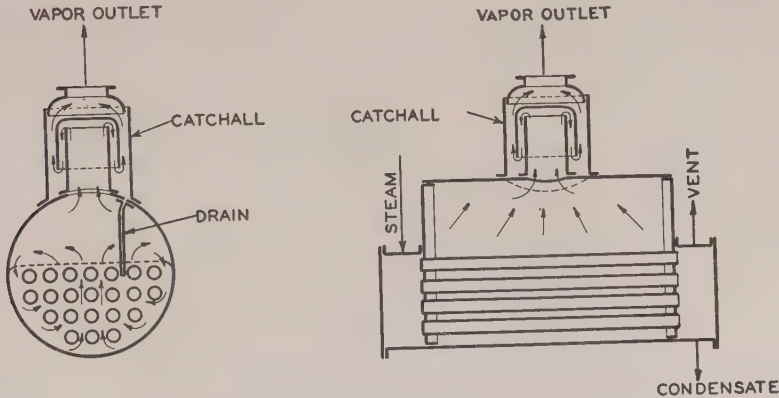


FIG. 1.—Sections of Old Rillieux Evaporator Operating in Cuba.

Some of the original Rillieux evaporators are still doing service in Cuba to-day, rebuilt and slightly modified, but still known as Rillieux effects. The bank of tubes assumed in a general way the shape of the shell, as shown in the sketch, extending perhaps from the mid diameter to a point near the lower part of the shell. The tubes were expanded in the tube heads without provision for expansion. Steam entered one end, and the condensate and gases were removed from the other, the tubes being pitched to provide natural gravity drainage.

Some of the old Rillieux evaporators found in Louisiana had different construction, having much larger tubes, expanded in one head, and blanked off on the other end, which was merely supported allowing for free horizontal variation due to heat changes. The tubes were pitched towards

the front head, and were perhaps 3 inches diameter by 12 to 15 feet long. There was a false head in front of the tube sheet into which were

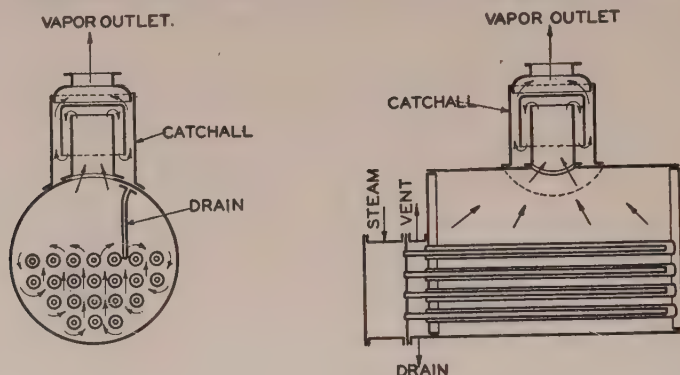


FIG. 2.—Sections of Rillieux Evaporator Operating in Louisiana.

attached smaller tubes, of the order of $1\frac{1}{4}$ inches diameter, each 3-inch tube having its corresponding $1\frac{1}{4}$ -inch tube threading into it, and extending



FIG. 3.—Standard Swenson Horizontal Tube Type.

to within a few inches of the blanked end. Steam was admitted into a chamber ahead of this false head, and entered each $1\frac{1}{4}$ -inch tube being

discharged into the 3-inch tube at its far end, thus steam had to flow backwards in an annular space in each tube towards the front head during condensation, both condensate and gases being removed from the space between the false head and the tube sheet proper.

One of this type of evaporators was still in use in Louisiana about five years ago. The design no doubt possesses merit, and doubtless gave very good results, although somewhat expensive in first cost. Figs. 1 and 2 show both designs. It is easy to see how the present horizontal tube evaporators grew from the original.

The Swenson effect (Fig. 3), which has seen many applications is an outgrowth of the type first mentioned above. The shell has been changed from the cylindrical shape of the original to a rectangular form with a semi-cylinder on top in order to provide more vapor space above the

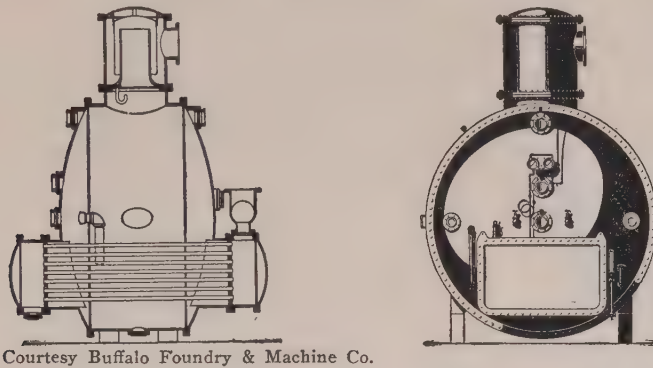
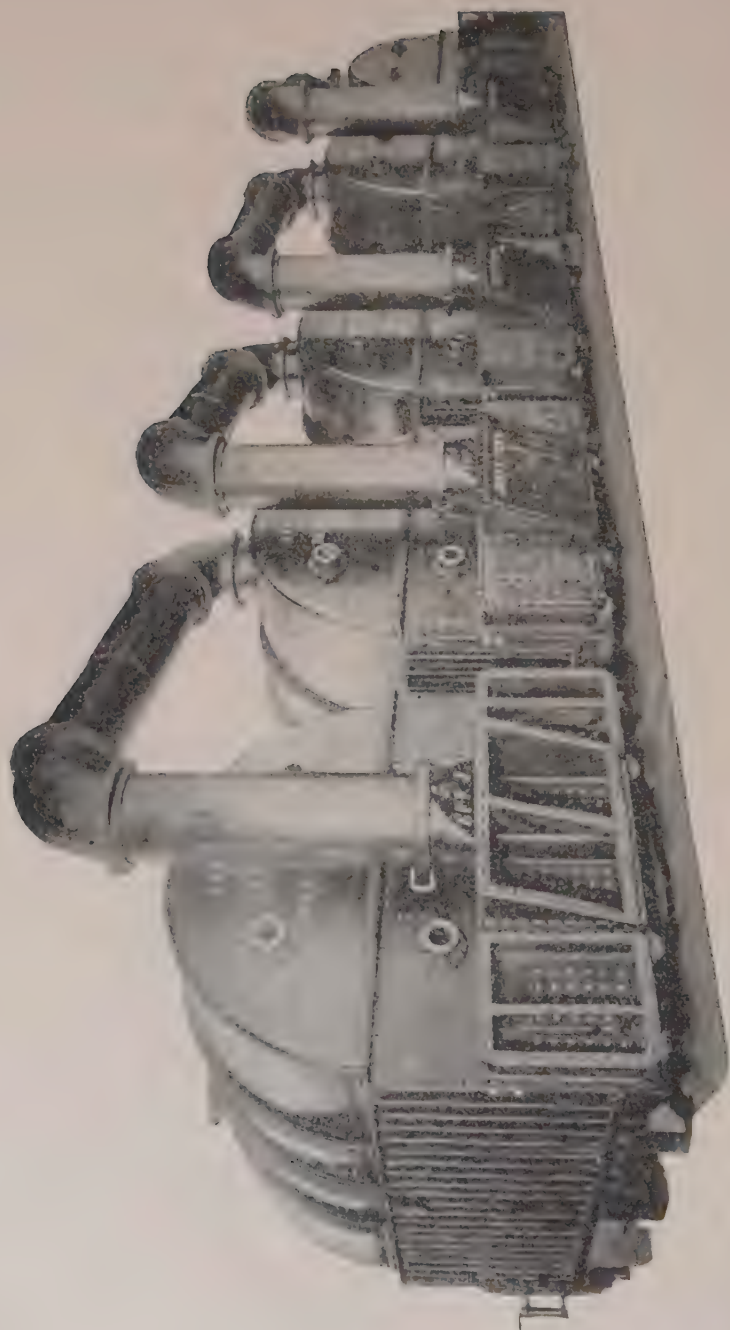


FIG. 4.—Horizontal Tube Evaporator.

boiling surface. The tube sizes have been changed from 2 inches to $1\frac{1}{4}$ inches, and instead of attaching them to the tube sheets by expanding, they are all held in stuffing boxes in both heads. The tube bank is perfectly rectangular, the bottom is not a perfect semi-cylinder, but a segment of a cylinder. The flat surfaces have to be braced against collapsing pressure by ribbing and stay bolts with separators. Steam only makes one pass through the surface. (Refer to chapter on Types of Steam Belts.) In the same family may be classed the familiar Zarembo evaporator discussed elsewhere in detail.

The Buffalo Horizontal Tube Evaporator (Fig. 4) is in reality a Rillieux apparatus, in which the tube bank is rectangular, providing to a great extent similar down comer features to those of the Zarembo apparatus. It has a horizontal cylindrical shell, as the original Rillieux, but the diameter is much larger in proportion to the tube bank than that of the latter. It is also a single pass apparatus.

The Wellner-Jellinek Evaporator (Fig. 5) is very similar to the Swenson apparatus, except that the steam makes three passes through the tube



Courtesy Kilby Manufacturing Co.

FIG. 5.—The Welther-Jelinek Evaporator.

bank as outlined in the chapter on Types of Steam Belts. This apparatus also has been very successful.

As regards the general character of the horizontal tube evaporator, the following might be pointed out:

1. It is true that the height of the tube bank is less than the height of the tubes in the vertical tube evaporator, and its partisans claim that it therefore suffers less from hydrostatic head losses.

2. The tubes being placed in stuffing boxes in both tube sheets are easily removable for cleaning or renewal.

3. It is practically impossible to scrape the tubes when there is obstinate scaling.

4. In view of the horizontal position of the tubes, it is probable that this type suffers more from blanketing of the heating surface by condensate than the vertical tube effect.

5. For conditions when there is fouling on the steam side of the tubes (such as in some cases in evaporating pyroligneous acid), there is quite an advantage in having mechanical access for cleaning, which would be impossible with vertical tube apparatus. This eventuality only occurs very rarely, however, and therefore the advantage is limited in its application.

6. The area of the boiling surface in proportion to the heating surface is probably larger than for vertical tube apparatus, and this evaporator is therefore unsuited to the handling of foamy solutions. On the other hand, it is possible that the boiling temperature rise of the solution is less than for vertical tube apparatus.

7. Usually, in first cost per square foot of heating surface it is less expensive than the other type, particularly for small and moderate sizes.

8. As to the relative merits for transmitting heat, in other words, efficiency of heating surface, as compared with vertical tube apparatus, authorities differ, and there is much to be said. The preference may be for one or the other according to local practice and precedent. It is a fact that in the cane sugar industry in Cuba, there are very few horizontal tube effects in use, and many vertical tubes. In the beet sugar industry, the reverse is the case. It is also a fact that cane sugar juices foul tubes much more rapidly than those from sugar beets, and that in the cane sugar industry, it is necessary to clean every week, whereas with beet sugar, the cleaning periods are much further apart, perhaps once or twice a campaign. This, no doubt, has a considerable influence on the preference, for when mechanical cleaning is necessary, it can be readily done with vertical tubes, and is almost impossible without dismantling the tubes of the horizontal apparatus.

Laboratory tests of horizontal tube evaporators as compared with vertical apparatus, as far as the authors know, show the advantage decidedly in favor of the vertical tube design. Mr. Manoury, the famous European authority, makes the statement that with equal conditions, the two types are equally efficient as far as heat transmission is concerned. It may be prejudice on our part, but the data available to us does not justify such a conclusion. The reasons are as follows:

1. It is much easier thoroughly to drain the water from the surfaces of a vertical tube evaporator, and its tubes will not be blanked off with condensate nearly as much as with the horizontal design.

2. It is very difficult properly to vent all the tubes of a horizontal evaporator, particularly the one pass design, in which the velocities of steam travel dwindle down to zero at the outlet end of the tubes.

3. It is impossible to maintain a uniform velocity of steam travel through the heating surface, thereby ensuring a thorough scrubbing of the

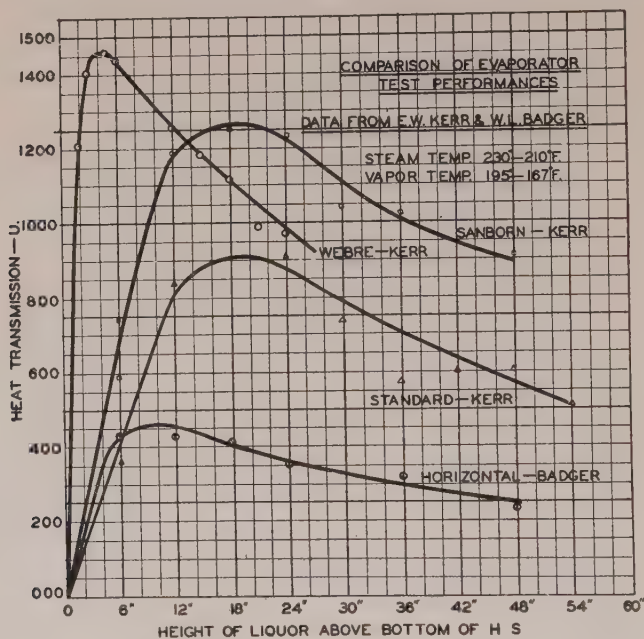


FIG. 6.

residual gas film, as is so easily accomplished with the vertical design, especially if the calandrias are properly baffled to suit operating conditions.

4. Due to the much smaller cross section for the path of vapor and liquor through the surfaces, the velocity of this mixture or, in other words, the circulation through these will be greater with the vertical than the horizontal evaporator. Inasmuch as it is generally admitted that the velocity of circulation on the liquor side is one of the prime determining factors of heat transmission, the vertical tube apparatus should show a higher heat transfer per square foot of heating surface.

5. As to static head required to produce proper circulation, it is very doubtful if the horizontal tube evaporator takes any less than a properly designed and operated vertical effect.

6. No doubt, in accordance with the principles brought out in the chapter on Natural Circulation, the velocity of liquor circulation will be greater in the horizontal tube effect, but it is undoubtedly true that the average for vapor and liquor will be much more for the vertical tube apparatus, and this average is the important determining one.

7. Very careful tests conducted by Professor W. L. Badger at the University of Michigan on Horizontal Apparatus, of which he is an exponent, compared with similar tests by E. W. Kerr, Efficiency Engineer of the Cuba Cane Sugar Corp. on vertical tube evaporators, show a decided advantage for the latter, the data from some of these tests being reproduced in Fig. 6.

8. It has been attempted to present the facts without prejudice, and the reader is therefore left to draw his own conclusions. It is believed that there is a field of application for each type, the vertical design having the advantage.

Chapter 5.

The Lillie Evaporator.

The Lillie Evaporator, which has been fully described in the current literature, deserves special mention, as its design displays unusual originality, and shows both operating and structural characteristics quite different from other designs. (See Figs. 1 and 2.)

Essentially, it is a horizontal evaporator with one tube sheet. The shell is cylindrical, and is provided with a full diameter door at the steam inlet end, as shown in the illustration. The tube sheet, which extends the full diameter, is secured to the shell a short distance from the door, the

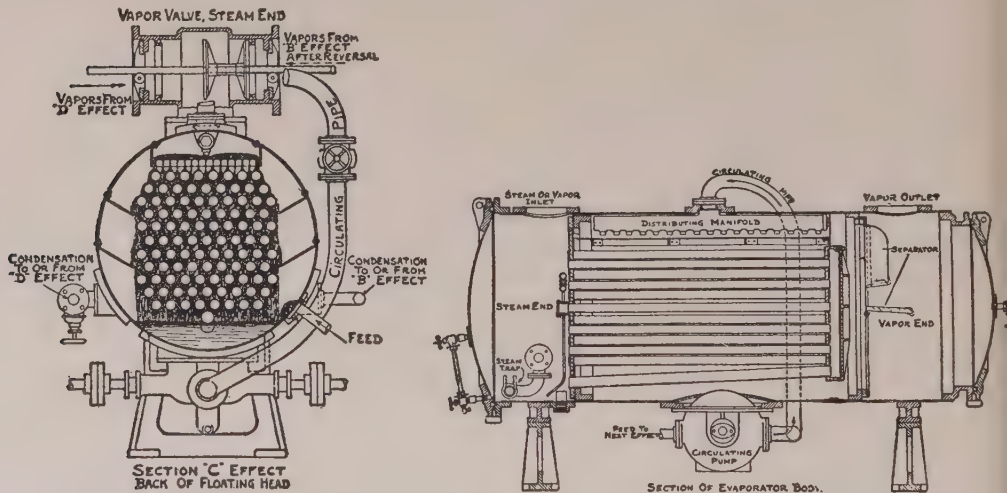


FIG. 1.

the intervening space forming a steam admission reservoir into which steam or vapor is admitted.

The tubes are expanded into this tube sheet, which is drilled in such a way that the tubes will all pitch towards the steam inlet end. They are of relatively large diameters and short lengths, perhaps $3\frac{1}{2}$ inches diameter, and 72 inches long, and have their far ends unsupported, and plugged, with a small vent orifice in each, giving individual outlets for non-condensables into the liquor space. The condensate flows towards the tube sheet or front end in the opposite direction to the entering steam, and

drops into the bottom of the steam space whence it leaves by the outlet for this purpose.

The provisions for expansion and contraction are therefore ideal, although the flat tube sheet must be designed to withstand collapsing pressure.

The liquid being evaporated is mechanically pumped from the bottom of the evaporator into a distributing tray above the tubes, so that it splashes down over the bank of heating surface in the form of thin sheets. The level of the liquor in the bottom is maintained by a float control in a small

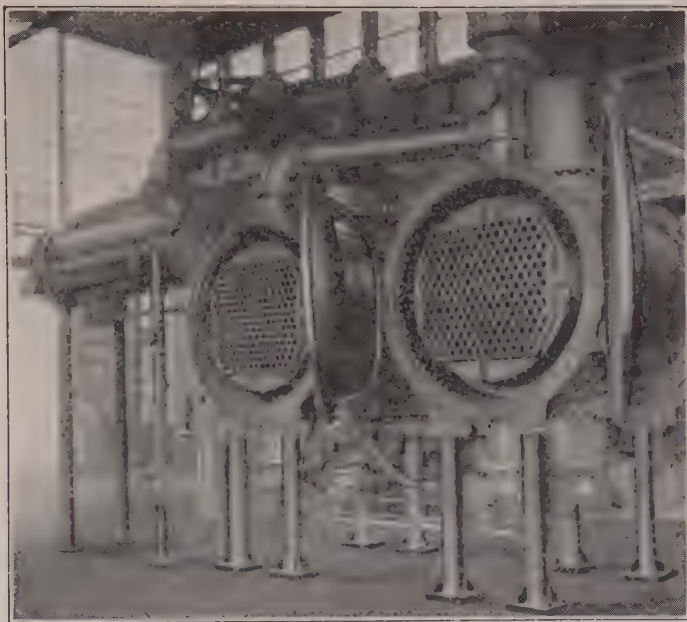


FIG. 2.

chamber attached to the shell below, and thus only the spray reaches the tubes, and no body of liquor. Thus a true film evaporator is obtained, and there is no "static head." The liquor circulation under the conditions is governed by the operation of the pump only, and is unaffected by any other cause.

The flow of feed is usually parallel current, *i.e.*, in the same direction as the flow of steam, although there is no reason why some other scheme cannot be used if found advantageous.

When operated as a multiple effect, it is characteristic of the Lillie design to have two condensers, or one condenser with vapor connections from the first and the last effects. The vapor, liquor, and condensate piping is so arranged that it is possible quickly to reverse the direction of flow of all

of these, changing the first effect into the last effect, and the last into the first. Such reversals of flow change the concentrations and temperature strains throughout, and it is claimed by the designer that this method prevents the formation of scale on the heating surfaces.

The performance of the apparatus in certain industries has shown the claims to be well founded, although it has not proven infallible for all solutions.

No doubt, the amount of liquor in transit in the apparatus at one time is extremely small, and, therefore, it is well adapted to the evaporation of delicate materials adversely affected by heat.

The heat transmission of Lillie evaporators is very good, and permits the use of many bodies in series. This design has been used with signal success in the production of distilled water on ship board, and in this work it has displayed remarkable freedom from incrustations. The objectionable features are the complications incident to mechanical circulation, and the rather elaborate system of piping for vapor, feed, and condensate.

Chapter 6.

"Buflovak" Inclined Rapid Circulation Evaporator.

(Courtesy Buffalo Machine & Foundry Co.)

The outstanding feature of the "Buflovak" Inclined Rapid Circulation Evaporator is that it can be kept sanitary and can be cleaned easily and quickly. This feature makes it particularly suited to handling food products; such as whole milk, skim milk, buttermilk, whey, beef extract, grape juice, cider, coffee, coffee substitute, gelatin and similar products.

This apparatus also offers the advantages of a small quantity of material in process, low hydrostatic head and rapid circulation of the liquid, with its consequent high rate of evaporation. Foam is broken up in the tubes and by the impact of the liquor striking the baffle plate, thereby reducing danger of entrainment.

The combination of these features makes the evaporator suitable for concentrating such products as glue, medicinal extracts, vegetable glue, animal and fish glue, saponification glycerine, syrup, wool scouring liquor, pyroligneous acid, black liquor, tanning and dye-wood extracts.

In extraction processes, using volatile solvents, the dilute extract can be passed through the evaporator, and the major part of the solvent rapidly distilled off and returned to the extractors. For this purpose the apparatus is operated under atmospheric pressure and equipped with a surface condenser.

Where viscosity permits, concentrations of 60 to 70 per cent. can be obtained without difficulty.

Careful tests,* with distilled water, tabulated below, show the heat transmission that may be obtained and the difference caused by steam entering either at the bottom or the top of the steamchest. Variation in

TABLE I—STEAM ENTERING BOTTOM OF STEAM CHEST.

Test	Steam Pressure Lbs. Ab.	Vacuum Inches Mercury	Steam Temp. F.	Boiling Point F.	Temp. Diff. F.	Feed Temp. F.	Heat trans- mission Coefficient
							B.t.u. /sq. ft. /hr/F.
1	7.35	26.85	170	116	63	54	620
2	9.8	26.25	192	122	70	70	680
3	9.8	26.25	192	122	70	58	695
4	12.25	25.6	203	128	75	65	780
5	14.7	24.75	212	135	77	66	885

* Van Marle, *J. I. E. C.*, 16, p. 452.

level of the liquor and in temperature of the feed liquor do not have a very great influence on the heat transmission. Table III shows tests made at a temperature of the liquor of about 140° F., and indicates that the high velocity of the liquor has a very large influence on the heat transmission. The heat transmission greatly increases with increasing temperature difference and is no longer proportional thereto.

TABLE II—STEAM ENTERING TOP OF STEAM CHEST.

Test	Steam Pressure Lbs. Ab.	Vacuum Inches Mercury	Steam Temp. F.	Boiling Point F.	Temp. Diff. F.	Feed Temp. F.	Heat Trans- mission Coefficient
							B.t.u. /sq. ft. /hr/F.
6	7.35	26.55	179	119	60	61	630
7	9.85	26.1	192.5	123.5	69	63	720
8	9.55	25.55	191	128.5	62.5	76	835
9	12.4	25.0	203.5	133	70.5	62	900
10	14.7	24.3	212	138	74	62	1,020
11	16.95	23.95	219	140.5	78.5	75	1,060
12	18.5	23.55	224	143	81	75	1,140
13 ¹	9.75	26.2	192	123	69	63	680
14 ²	9.6	26.25	191.5	122	69.5	63	660
15 ³	12.25	24.6	203	136	67	62	960
16 ⁴	12.25	24.4	203	137.5	65.5	62	995

¹ 12-in. level.² 30-in. level³ 18-in. level.⁴ 27-in. level.

TABLE III—STEAM ENTERING TOP OF STEAM CHEST.

TEMPERATURE LEVEL APPROXIMATELY 140° F.

Test	Steam Pressure Lbs. Ab.	Vacuum Inches Mercury	Steam Temp. F.	Boiling Point F.	Temp. Diff. F.	Feed Temp. F.	Heat Trans- mission Coefficient
							B.t.u. /sq. ft. /hr/F.
17	8.25	23.75	184	142	42	51	465
18	9.7	23.75	192	142	50	59	835
19	11.8	23.75	201	142	59	69	890
10	14.7	24.3	212	138	74	62	1,020
12	18.5	23.55	224	143	81	75	1,140

Other tests compare the performance of this evaporator with that of the standard horizontal and vertical type for waste liquor from paper mills. Influence of the viscosity of the liquor can be clearly seen by comparing tests C and D and show the advantages of staggered flow of the liquor.

The heat transmission coefficient is practically the same for 52 per cent. liquor under 11 inches vacuum and 43.5 per cent. liquor under 25.5 inches vacuum. With staggered flow a higher concentration can be obtained without reducing the rate of evaporation.

TABLE IV. TESTS WITH WASTE PAPER MILL LIQUOR.

Test	A. Hor. Tube	B. Vert. Tube	C. Incl.	D. Rapid Circulation	E.
Type Evaporator					
Density, B at 130° to 140° F.	30	30	30	32	33
Solids, percentage	46.88	44.44	43.48	52.05	55.72
Steam pressure, lbs. gauge.....	5.45	5.9	4.9	4.75	5.15
Vacuum, inches, Hg.	25.91	25.34	25.68	10.9	11.1
Liquor level, inches	13	23	26	42.5	43.5
Feed temperature, F.	68	94	70	112	101
Steam temperature, F.	228	229.5	227	226.5	227.5
Liquor temperature, F.	133	130	135	198.5	199
Temperature difference, F.	95	99.5	91	28	28.5
Evaporation, lbs./hr./sq. ft.	15.40	20.70	27.30	7.72	4.84
Evaporation, lbs./hr./sq. ft., F.	0.162	0.208	0.300	0.276	0.170
Heat transmission, B.t.u./hr./sq. ft./F.	175	219	324	294	183

Construction. This type of evaporator is constructed as shown in Figs. 1 and 2 with an inclined steam chest of comparatively small diameter, fitted with tubes of considerably greater length than are used in the conventional type of evaporator. The liquor boils inside the tubes, the steam is on the outside.

The downtake is entirely separate from the steam chest. This permits a free flow of the returning liquor from the vapor body to the bottom section, and is a great advantage over a downtake, built in the steam chest and surrounded by steam. Such design retards the flow of the returning liquor and often causes foaming. The steam chest and downtake are rigidly bolted, by a flanged connection, to the conical bottom of the vapor body. The other ends are connected by means of the bottom section. This section has a threefold purpose. It forms a passage for the return of the liquor from the downtake to the steam chest. The bottom section is provided with a door and an extension. The door affords access to the evaporator tubes and downtake. The extension is constructed to rest on a suitable pier, thus carrying part of the weight of both steam chest and downtake.

The tubes are expanded in the tube plates, making a steam-tight joint without the use of gaskets or packing, which often become so troublesome in operation. The steam inlet to the steam chest is placed to obtain the maximum heating effect from the steam and to sweep the non-condensable gases to the opposite end. The outlet drain is installed at the lowest point to remove all condensation thoroughly and rapidly.

In general practice, the steam inlet is placed at the top of the steam chest, so that the flow of steam over the tubes will sweep the condensate down. For milk and similar products, which tend to bake on the tubes, better results are obtained by placing the steam inlet at the bottom, as shown in the illustration. Although this arrangement reduces heat transmission to a certain extent, it largely overcomes the tendency of the milk to bake on the tubes.

The vapor body, of comparatively small area, is usually made of cast

iron, or in the case of the sanitary type used for food products, it is made of copper. This part, like the tubes in the steam chest and the downtake, may be made of any other material, best suited to the nature of the liquor.

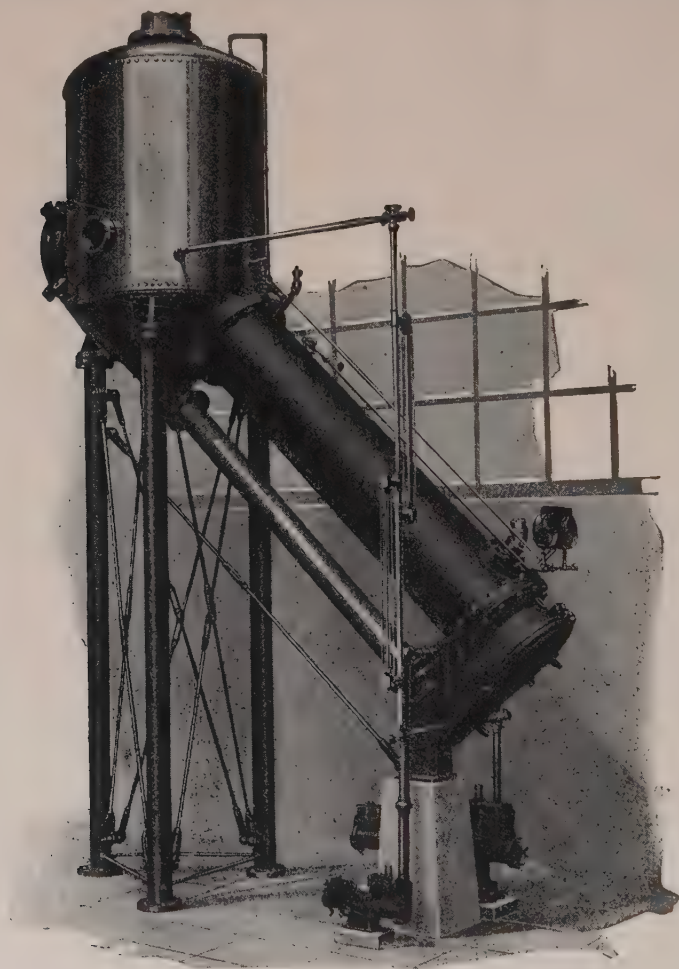


FIG. 1.

Tinned copper, aluminum, Monel metal or enamelled steel are other materials that may be used without making the evaporator too expensive.

The construction of vapor body and steam chest as separate parts makes it possible to design each in the most economical manner. The diameter of the steam chest is determined by the number and size of the tubes re-

quired to make up the heating surface. The diameter and height of the vapor body only needs to be large enough to carry the vapor without entrainment. In multiple effect evaporators of this type, advantage can be taken of these facts to reduce the size of the vapor body of each effect

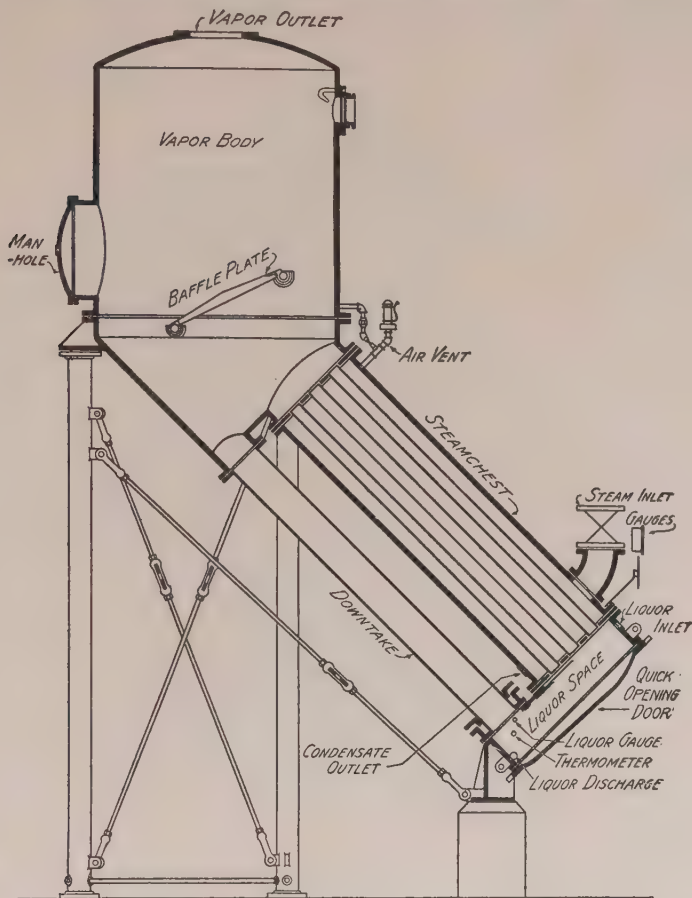


FIG. 2.

in accordance with the quantity and volume of vapor produced. A material reduction in first cost is effected in this way.

The upper part of the evaporator is supported by three heavy cast iron columns with substantial cross braces and spacing rods at the bottom. This construction makes the evaporator self-supporting and extremely rigid.

A hinged door, which is held tightly in place by swinging bolts, is

provided to give access to the tubes in the steam chest and to all parts of the vapor body.

Observation glasses are conveniently located, so that the action of the boiling liquor may be observed. Means for cleaning the inside of these glasses have been provided.

A baffle plate is placed in the vapor dome at the proper angle so that

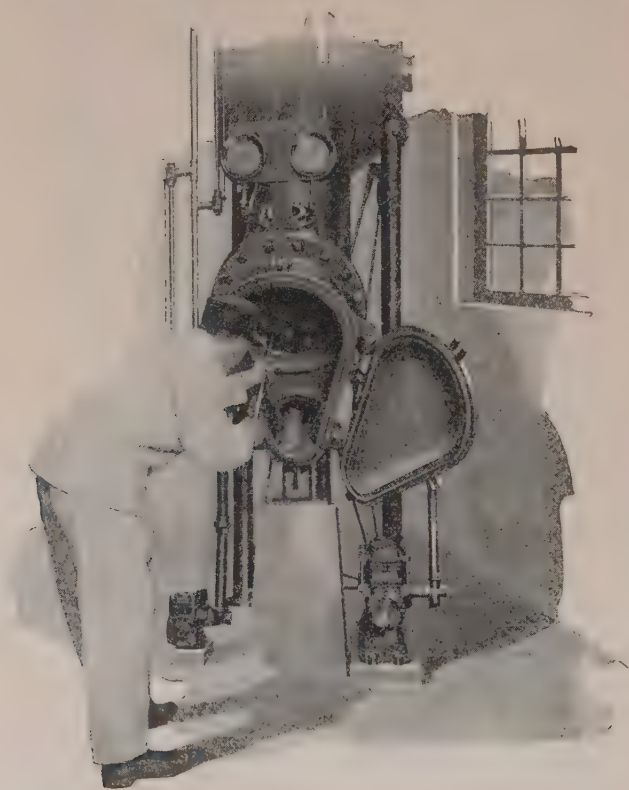


FIG. 3.

the liquors, which are thrown against it as they leave the tubes, are deflected directly into the conical bottom and returned through the downtake.

Operation. The liquor to be evaporated is admitted at the top of the bottom section, inside the tubes, to be discharged at high velocity against the baffle plate in the vapor dome. In striking the baffle plate, vapor and liquor are separated, the liquor being deflected, so that it falls directly into the conical bottom of the vapor body, flows down through the downtake and is recirculated.

The evaporator can be operated continuously in the usual manner. When the liquor reaches the required density, the liquor pump is started

and allowed to withdraw the concentrated liquor continuously from the lowest point in the bottom section. By regulating the speed of the pump, the concentration can be maintained at the desired density.

The best method for handling large batches is to connect a storage tank with the evaporator and to draw the liquor from the tank into the evaporator. In this method, the liquor is continuously circulated between the evaporator and the storage tank until the required density is reached. This process has the advantage of maintaining a high rate of evaporation in cases where high concentration of the liquor or uniform density is required. This is due to the fact that the material is evaporated at its highest density for a comparatively short period of time; and only at the finish of the batch. The greater part of the evaporation is accomplished at the lower concentration. In most cases this enables the evaporator to be run at a much higher capacity than if it were run continuously.

The ease with which the evaporator may be cleaned is shown in Fig. 3. Inaccessible nooks and corners where materials might lodge and places difficult to clean, have been carefully avoided, both in design and construction.

Cleaning the evaporator tubes, which is such a laborious task in the ordinary type of evaporator, is very easily accomplished by opening a hinged door at the bottom. This door also exposes the downtake tube.

The manhole is sufficiently large to allow the baffle plate to be removed either for cleaning or inspection. With the baffle removed, the smooth walls of the vapor body may be thoroughly and quickly cleaned. Careful thought has been given to the construction, so that no seams, corners or inaccessible places exist where materials are apt to lodge, or where it is difficult to clean thoroughly.

Chapter 7.

The Kestner Evaporator.

(Courtesy Kestner Evaporator Co.)

Five things are essential in a good evaporator, and the nearer they are approached, the nearer perfect the apparatus. These are: high velocity of the heating steam over the heating surface; high velocity of the liquid to be concentrated over the heating surface, the absolute elimination of non-condensable gases from the heating surface; the liquid in the evaporator the shortest possible time; no hydrostatic head.

These conditions mean high rate of heat transfer, complete use of the heating surface, and freedom from injury to the most sensitive materials. The film evaporator embodies the attempt to meet these conditions.

The Kestner Film Evaporator came into existence in the attempt to overcome the difficulty of obtaining an even distribution of the liquor to the tubes of horizontal film evaporators. The introduction of contracted orifices, spray mechanism, etc., in the latter brought many difficulties without producing the best of results.

There occurred to Paul Kestner, the internationally known French engineer, the idea of a nest of vertical tubes with their lower ends in a barrel, giving absolutely even distribution to all the tubes. Thus originated the Kestner Film Evaporator—a true film evaporator without any complicated devices for circulating, regulating, and distributing the liquid to be concentrated.

Long and careful study determined the exact tube dimensions necessary to produce a true film. (Many so-called film evaporators do not film at all.) These dimensions differ for different materials and operating conditions, and each apparatus is specially designed for the work in hand.

Further study was required to devise a method of catching the emulsion discharging from the tubes at a velocity of 90 to 150 feet per second, and to separate the liquid from the steam. This resulted in the Kestner patented separator which absolutely eliminates every bit of entrainment of the non-volatile matter.

In construction the Kestner consists (Fig. 1) of a long, narrow calandria with tubes 16 to 92 feet long—in one to four sections—depending upon the kind of work required. The tubes are expanded at their ends into tube sheets. This eliminates stuffing boxes with their constant attention and leaks.

On top of the calandria is the vapor belt in which are located the Kestner centrifugal separator and a catchall.

In the Climbing Film Type the tubes are 23 feet long. The lower ends of the tubes project into a feed chamber and an even feed of liquid to each tube is obtained by a simple and effective patented device. The mixture of concentrated liquor and vapor is discharged at high velocity from the upper ends of the tubes into a stationary centrifugal separator of special design, which imparts to it a rapid rotative movement, projecting the liquor against the walls of the vapor belt, while the vapor, being lighter, rises

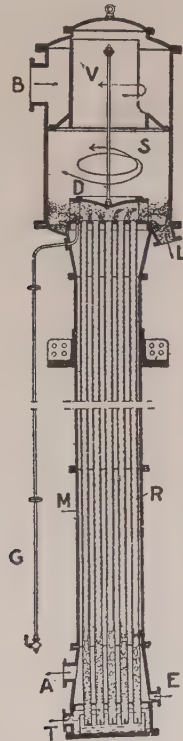


FIG. 1.—Kestner Vertical Single Effect Pressure Evaporator.

to the top and is drawn to the steam belt of the following effect or to the condenser, as the case may be. The liquid runs down the walls of the vapor belt into the dish-shaped bottom and passes down a discharge pipe into the feed chamber of the next effect or to the concentrated liquor tank, if from the last body.

Generally, the liquid to be concentrated is fed into the evaporator by a feed pump through a heater. If the liquid is supplied near the boiling point of the body, no heater is necessary. If it is fed cold directly into the pan, some of the heating surface must naturally be utilized to raise its temperature to the boiling point. If the liquid is heated by vapor

coming from evaporation in the body, the tubes will be filled solid for about three feet from the bottom. If the liquid is fed hotter, this height approaches zero.

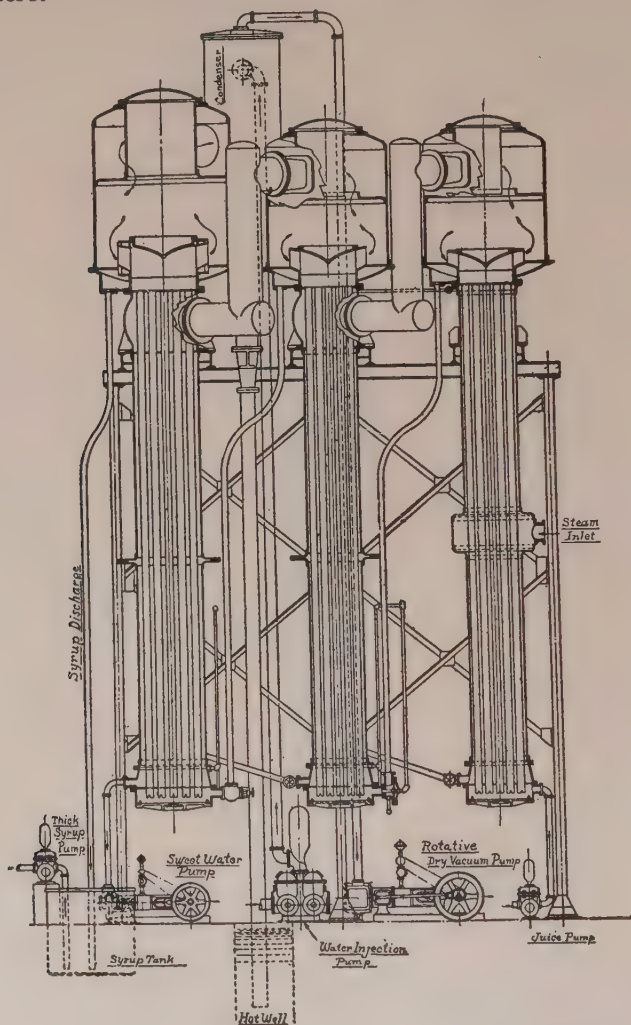


FIG. 2.

The moment the liquid starts to boil, the steam thus liberated rises rapidly, drawing a film of liquid up the sides of the tubes. Thus we have a perfect film, perfect distribution and great velocity over the heating surface. In fact, the liquid remains in contact with the hot surface in each

body not over one minute; once it has passed through the tubes there is no return, except in special cases.

The steam enters at the top of the long, narrow steam belt, driving the condensed water and the non-condensable gases rapidly to the bottom, where they are drawn off.



FIG. 3.—Kestner Triple Effect under Vacuum.

If the apparatus is a multiple effect as shown in Figs. 2 and 3 the liquid, after passing through the first body, goes to the bottom of the next, where it instantly flashes into a boil and is entrained up the tubes.

The entire operation is controlled by the feed valve; and the operator has no way of disturbing the level of the liquid in any pan. In fact, there is no volume of liquid in any body, the small feed chamber holding but a few gallons.

For viscous liquids, or where a very high concentration is desired in a single pass in a single pan, the falling film type is used. This is

arranged so that the liquor first enters "climbing film" tubes, the upper ends of which discharge into a small chamber. Here the liquor enters the upper ends of the "falling film" tubes, passing in a complete film down-

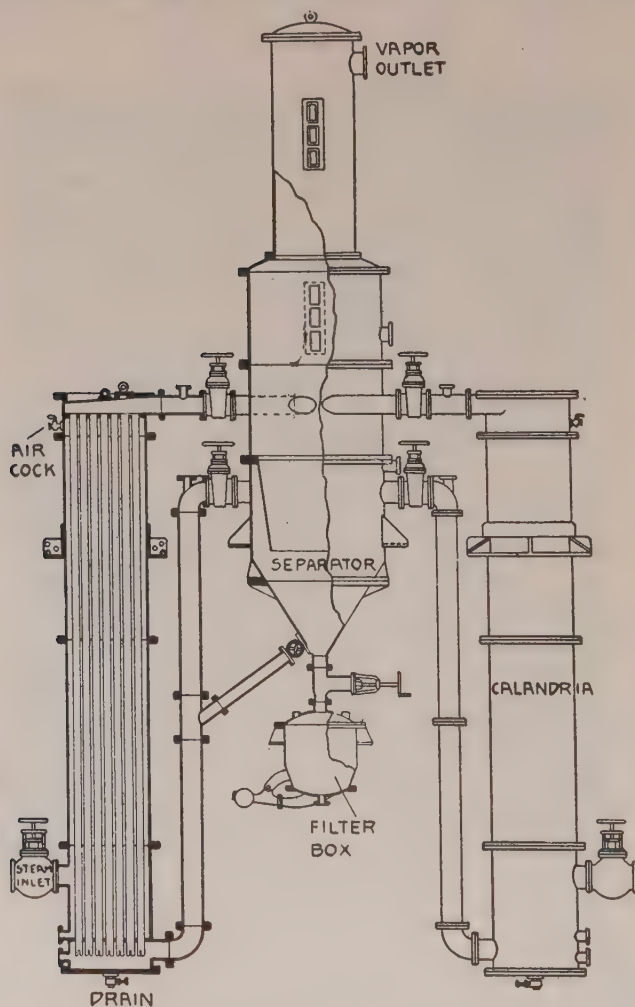


FIG. 4.—Kestner Salting Pan.

wards from the top to the bottom of the tubes. The separator in this case is placed at the bottom. This type is used with liquids that are so thick they cannot climb.

A special type (Fig. 4) is made for salting. This pan circulates the liquid through a long tube calandria discharging into a special vapor belt

where the separation of the steam and liquid takes place, and then the salt is separated from the liquid before the liquid returns to the calandria. There are all the advantages of the climbing film type.

These evaporators are also designed to be fire heated, using direct fire or spent gases. They are also operating successfully on subterranean gases.

The advantages obtained are:

1. Absolute extraction of the non-condensable gases.
2. Absolute elimination of entrainment.
3. Complete use of the heat of self-evaporation in the liquid and condense water.
4. Elimination of hydrostatic head. There is no body of liquid in the tubes of any body but the first. Hence all the surface is available for evaporation, and the apparatus will operate on the minimum temperature fall.
5. No volume of liquid in the apparatus.
6. And most important, the short time required for concentration. In a triple effect the entire concentration is performed in 3 minutes, in a quadruple effect, in 4 minutes. This permits handling the most sensitive materials at a higher temperature without danger of injury to the material.

Chapter 8.

The Badger High Speed Evaporator.

(Courtesy E. B. Badger & Sons Co.)

The Badger High Speed Evaporator (Fig. 1) was developed primarily to concentrate liquors of certain definite characteristics, the most important of these being liquors of a foamy nature or those harmed by long contact with the heating surfaces. There are other definite advantages to be gained by the use of this equipment, however, such as a high rate of heat transfer, lack of entrainment, complete removal of non-condensable gases, and ease of control.

Fig. 2 shows the general features of this machine. It is of vertical tubular type with tubes approximately 16 feet long, varying in diameter according to requirements. The tubes are encased in a vertical cylindrical steam belt, steam entering near the top and condensed steam and non-condensable gases being removed at the bottom. The body is flared near the upper portion, providing a support for the equipment. The tubes are expanded at the top and bottom into tube sheets; the upper tube sheet is affixed to the shell by diaphragm, providing necessary movement for difference in expansion between the tubes and shell.

Below the lower tube sheet is a liquor chamber provided with a drain at the bottom. Above the upper tube sheet is the vapor chamber, provided with customary sight-glasses, manhole and a removable top cover, cone shaped, to deflect the rising stream of vapor and liquor.

Large manholes of quick-opening type are supplied for ready access to the top and bottom chambers. The removable top on the evaporator proper and catchall provides ready access to these portions of the equipment for either cleaning or removal of tubes.

These evaporators may be constructed of any metals required, such as iron, steel, copper or bronze, depending on the liquors to be handled and their action on the various metals which might be used.

The units as shown in the cut may be connected up in any multiples required, such as double, quadruple or quintuple effect, depending on the conditions of the problem in hand.

This equipment is entirely non-foaming,—a very important feature. Foam results in losses by entrainment, and in the usual short tube type machine is almost impossible to prevent. Such excessively foaming materials as distillery slop, photographic gelatine and jack pine liquors have been handled in these machines with absolutely no foaming problem. No foreign materials, such as kerosene, are needed to control this factor.

Due to the efficient type of centrifugal separator used there is no entrainment whatsoever. When operated multiple effect, the condensate from the steam belts of these bodies is as clear as water and has been used for boiler purposes, washing pulp, etc.

There results from the long tubular design a high velocity on both the steam and liquor sides of the tubes. The high velocity of the liquor rising in the tubes causes a scouring action which maintains the tubes free from



FIG. 1.—Badger High-Speed Evaporator.

scale or coating and results also in the most optimum heat transfer conditions. High velocity on the steam side maintains the outside surfaces free from condensate, and from non-condensable gases. The steam belt of the equipment being constructed as a long, narrow cylinder, effects complete removal of non-condensable gases.

Positive circulation is provided by means of the outside separate down-take pipe, the circulating liquor being controlled by valve so as to maintain the most efficient conditions during the operation of the equipment. In other words, it is possible to supply in the bottom chamber of the evaporator just sufficient liquor to obtain a wetted surface at the top of the

tube without having an over supply which results in excessive static head loss, lower velocity, etc.

There are no mechanical or compressed air operated level controls, the liquor seeking its own level by the circulation control method. Compressed air level controls get out of order, plug up and cause shutdown of the equipment.

These units are simple and easy to operate; in one installation, four complete quadruple effect evaporators are operated by one man.

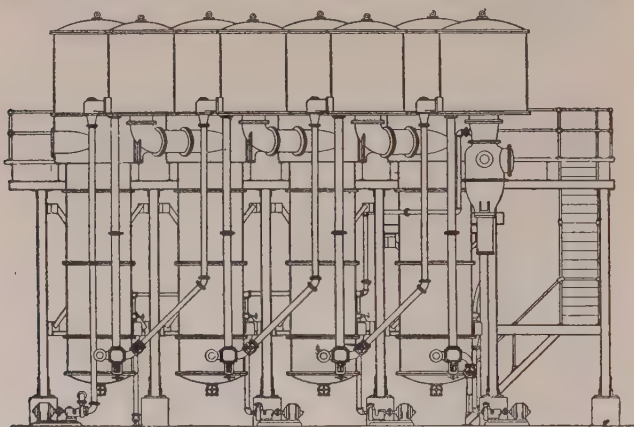


FIG. 2.—Badger-Webre High-Speed Evaporator.

Where liquors which are affected by long contact with heating surfaces have been used in this type of equipment, exceptional results have been obtained. With our high velocity of liquor up the tubes, exceedingly short time of contact results and it has also been possible to operate the units on a high vacuum with a partial vacuum on the steam side, thereby obtaining the most advantageous temperature conditions.

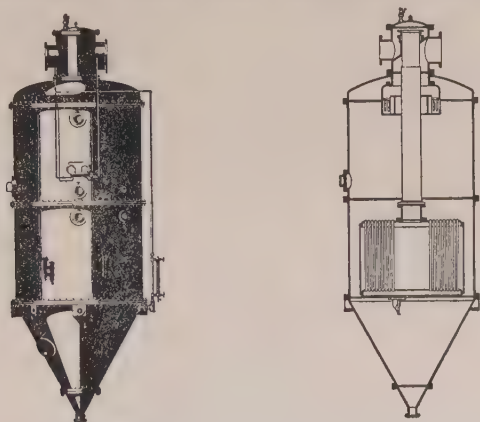
The following materials, among others, which are ordinarily difficult to handle in evaporators have been successfully concentrated in the Badger High Speed Type Evaporator: Distillery slop, photographic gelatine, maltose syrup, blood serum, paper mill waste liquors.

Chapter 9.

The Basket Type Evaporator.

The basket type evaporator, of which there are many forms, is designed to have its entire heating surface, as a unit, inserted into the inside of the evaporating body. Thus the heating element is completely surrounded by the liquid it is to evaporate. See Figs. 1 and 2.

This arrangement is very advantageous in permitting quick and easy repairs in case of necessity, and doubly so if there are a number of duplicate units in the same plant. Under these conditions, a spare basket is kept



Courtesy Buffalo Foundry & Machine Co.

FIGS. 1 AND 2.—“Buflovak” Basket Type Evaporator.

on hand, properly conditioned so that it can be available for replacement at a moment's notice.

In the Chemical Industries, this apparatus has found very useful applications in the concentration of crystallizing solutions, for it has a large, free annular downtake, permitting uninterrupted liquor circulation so essential in this class of work. The design in vogue is the familiar cylindrical basket with steam admission into the top head, as shown in the illustration.

Some designers make the bottom head of smaller diameter than the top, so as to take up the full space in the former with tubes at the same pitch. In addition the heads are both dished up, so that the tubes lean towards the outer periphery. This arrangement could not be better as

far as facility of circulation is concerned, and its popularity is a good testimony to its efficiency.

Occasionally, there is provided an inverted cone attached to the steam pipe slightly above the top tube sheet, whose object is to deflect the streams emitting from the tubes towards the outside and the annular downtake, and away from the direction of the vapor outlet at the top above, thus reducing the danger of entrainment.

The steam distribution is not of the best, and there is no doubt considerable mixing of condensables and non-condensables with reduction of

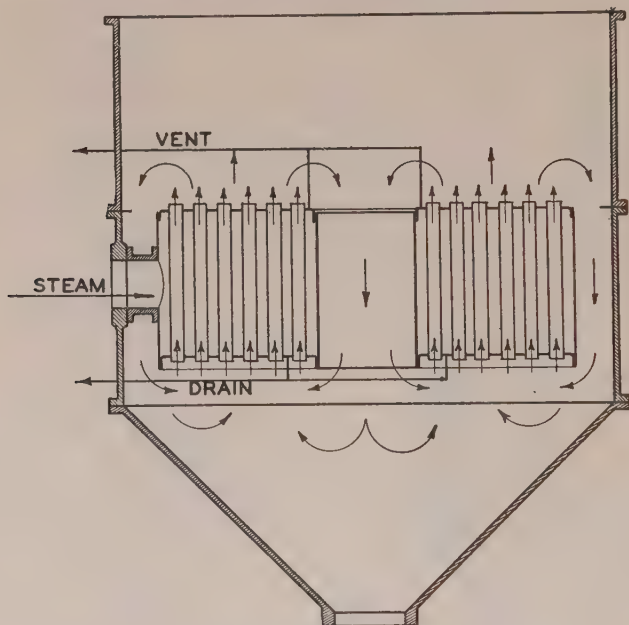


FIG. 3.—Sugar Vacuum Pan with Floating Calandria.

heat transmission. Some designers try to overcome this feature by inserting a disk immediately below the steam inlet on the inside of the basket, thus deflecting steam towards the outer diameter at the top, permitting the accumulation of non-condensables at the bottom in the centre for convenient removal.

When solutions are to be evaporated which are affected by heat, this design is unsuited without insulating the steam pipe above the basket, for drops of liquor are baked thereon and overheated.

In the sugar industry, this design is sometimes used, but steam is admitted at the side of the basket for the reason explained above. With such an arrangement, there is also provided a central downtake, and the surface can be properly baffled for steam distribution and gas removal.

Some sugar vacuum pans are built on this principle, sometimes with

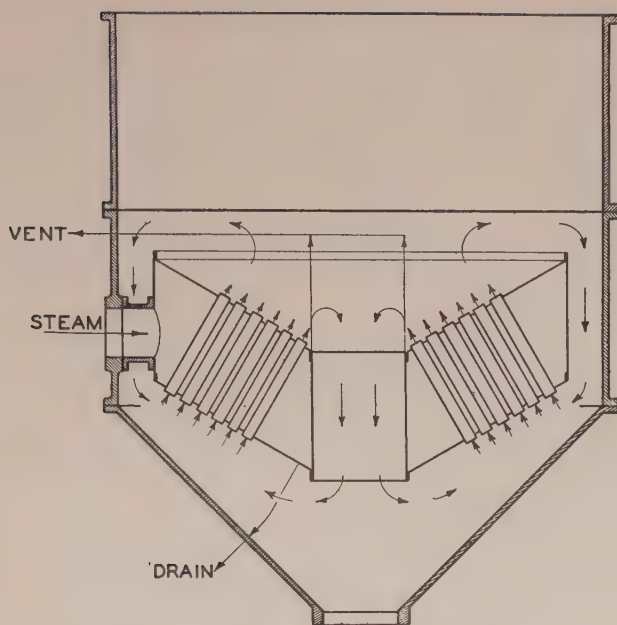


FIG. 4.—Sugar Vacuum Pan with Inclined Calandria.

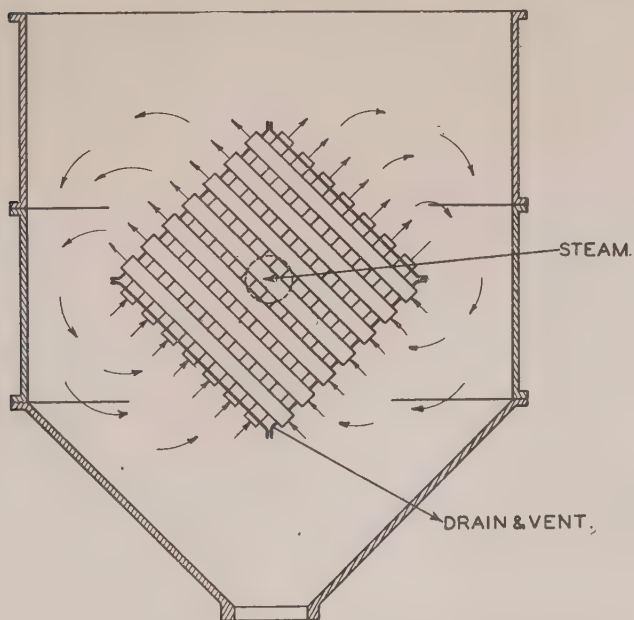


FIG. 5.—Special Evaporator with Box Heating Element.

flat calandrias (see Fig. 3), and other times with inclined or conical calandrias, such as the so-called "German express pan" (see Fig. 4).

Floating heating elements are also made of different shapes, one type being like a square box (Fig. 5), the four sides being tube sheets, the box resting on its edge, and the rows of tubes alternately pointing to the right and to the left.

In all these designs, facility of liquor circulation has been the prime consideration, and it is very important for crystallizing solutions as has been pointed out before.

There is a decided tendency in the Sugar Industry to abandon this type in favor of the stronger, simpler, and cheaper straight calandria pan, which works equally well and gives very little trouble.

When the solution is of a relatively non-viscous nature, and carries salts in suspension, there may be serious erosion at the upper parts of the tubes particularly when operating at high vacuum, where the velocities are unavoidably high.

Chapter 10.

The Garrigue Evaporator.

(Courtesy William Garrigue Co.)

The design of the Garrigue Evaporator is the result of continued efforts to combine scientific design with practical requirements and may be cited as an example of controlled circulation of the liquor in the evaporator.

The evaporator, as arranged for the evaporation of soap lyes for the recovery of glycerine and salt, is shown in cross section in Fig. 1. It consists essentially of three parts, namely, the calandria or heating element, the evaporator body or vapor belt which provides space for the vapors to separate from the boiling liquor, and the cone bottom in which the salt is deposited as it crystallizes out. The accessories to the evaporator are the catchall, which separates out any liquor mechanically carried over with the vapors; the condenser in which the vapors from the evaporator are condensed; the dry air pump which removes the non-condensable gases from the condenser; and the salt extractor which provides a means for intermittently removing the salt as it is deposited.

The calandria of the Garrigue Evaporator consists of a flanged ring, usually of cast iron, and closed on each end by heavy copper tube sheets. The upper portion of the calandria is flared out to the diameter of the evaporator body, thus making the upper tube sheet of larger diameter than the lower tube sheet. The heating elements are vertical seamless drawn copper tubes, the ends of which are expanded and beaded into the tube sheets. The larger diameter of the upper tube sheet allows sufficient flexibility to take care of the expansion of the tubes and thus allows a rigid connection to be made between the tube sheets and the tubes. In this manner it is possible to eliminate entirely the usual packings and glands at the points where the tubes pass through the tube sheets. Steam is admitted to the calandria through a suitable flanged opening in the side of the calandria ring. The condensate is drained off from the lowest point of the calandria.

Fig. 2 shows the calandrias for two 8-foot diameter evaporators each containing 1500 square feet of heating surface. It will be seen that the central portion of the tube sheet is fitted with small tubes, in this case 2 inches inside diameter, and the outer portion of the sheet is fitted with larger tubes, 4 inches inside diameter. The purpose of such an arrangement is for augmenting the circulation of the liquor in the calandria. In the center of the calandria the liquor should take an upward flow. This is brought about by a system of baffles (U. S. Pat. No. 1,298,925) which

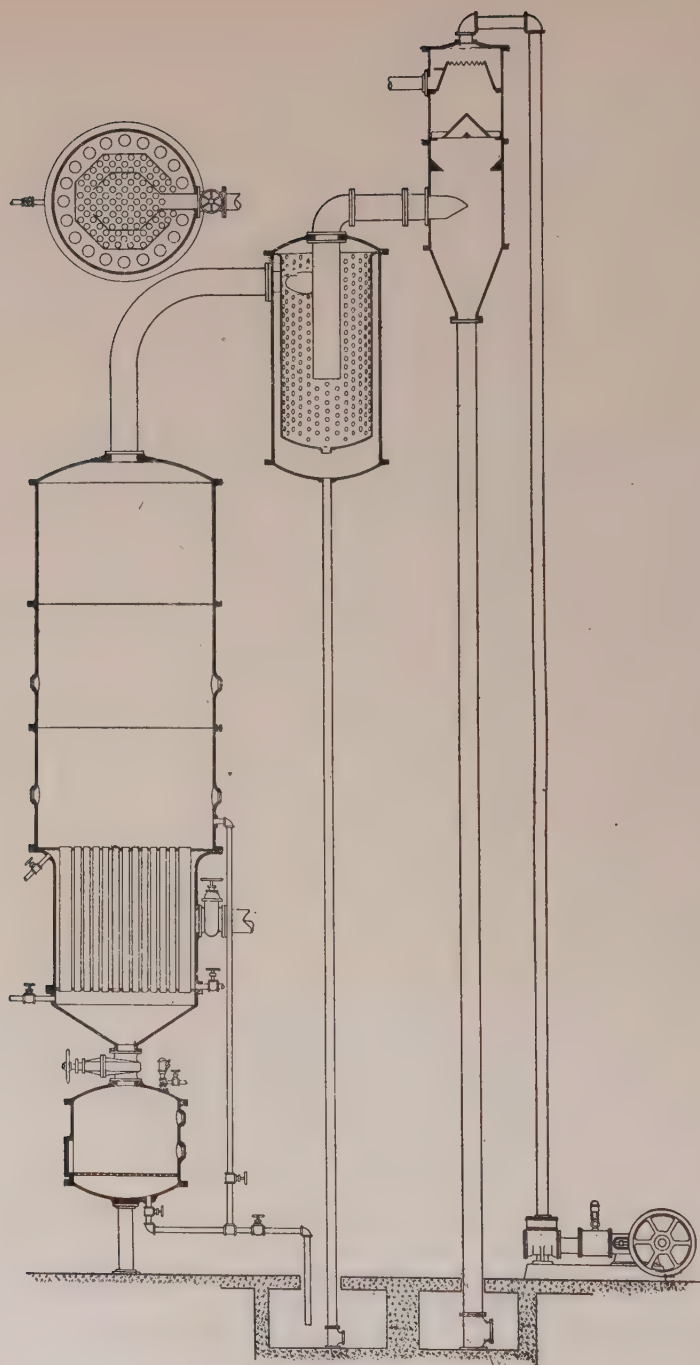


FIG. I.
448

will be described later and which carries the heating steam directly to the center of the calandria, giving the maximum heating effect at this point. By making these tubes small in diameter, high velocity and also high heat transfer are obtained. The larger outer tubes mentioned in the previous paragraph, together with the steam baffling arrangement, make it possible to maintain a rapid circulation and high heat transfer in the central tubes,

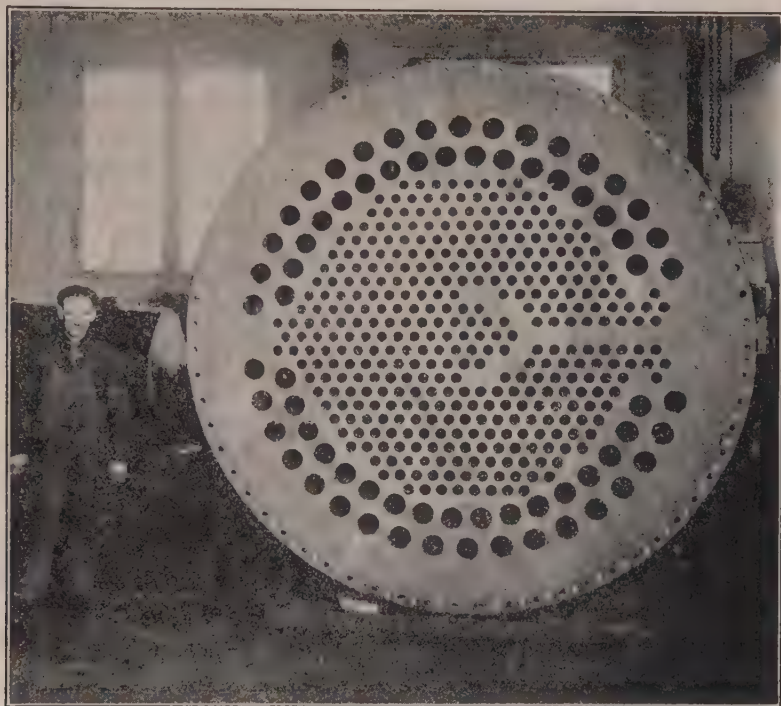


FIG. 2.

and a low heat transfer and slow downward flow in the "down-comer" or outer tubes.

The vapor belt of the evaporator is built up of flanged ring sections bolted together with asbestos gaskets between machined joints. The use of ring sections allows easy transportation and erection and at the same time provides for a minimum number of joints. The inside diameter of the evaporator body is larger than the diameter of the calandria in order to give the top tube sheet flexibility to take care of tube expansion and at the same time give room for the waves of liquor to flatten out. The purpose of the vapor belt is to provide a space wherein the vapors and boiling liquor can separate so that only vapors will leave the outlet at the top of the evaporator. The height of the vapor belt is dependent to a

great extent on the nature of the material which is to be handled in the evaporator. Liquors which show a tendency to foam during evaporation naturally require a considerable volume in which to effect this separation, and the vapor belt is consequently made higher. The vapor belt is provided with observation glasses in each ring, and to it are also attached the vacuum gauge, thermometer, liquor inlet, vacuum breaker, and liquor tester. The top of the vapor belt is closed by means of a domed cast iron cover provided in the center with a vapor connection leading to the catchall. To the bottom of the calandria is bolted a cast iron conical bottom connected by means of a valve to the salt extractor. The deep cone provides a suitable place for the salt to settle out and to flow through the valve into the salt extractor.

The catchall is an important accessory to the evaporator. Its purpose is to trap all entrained liquor which may be carried over from the evaporator with the vapors and to return it to the evaporator. The catchall is of circular cross section and usually built of cast iron. The vapors from the evaporator enter it tangentially in the side near the top, producing a whirling motion of the vapors which will throw any liquor to the sides of the catchall and allow it to drain out at the bottom. The vapors leave the catchall through a pipe entering the catchall at the top and extending about one-half way to the bottom. In the Garrigue Evaporator the catchall is drained by means of a barometric column (U. S. Pat. No. 1,317,488). This barometric column terminates at its lower end in an open hotwell or tank. The purpose of this is to provide a means of signalling the operator when an unusual amount of liquor is being carried over into the catchall. Also, in case of excessive frothing or boiling over, the excess liquor can be held in this hotwell or tank until normal conditions have again been established in the evaporator. The liquor is then gradually drawn back into the evaporator by means of a suitable pipe connection. This method of operation is, of course, not possible with the usual type of catchall which drains directly back into the evaporator and in which large losses can easily take place without being detected by the operator.

After passing through the catchall, the vapors pass into the condenser. The condensing equipment on the earlier type of evaporators usually consisted of a combined jet condenser and wet air pump. This system, while low in first cost, is inefficient for the reason that the pump not only has to handle the non-condensable gases but also has to handle all of the condensing water. A more efficient condensing plant consists of a barometric condenser operated in conjunction with a dry air pump of the power driven, low clearance type.

The general construction of the Garrigue Barometric Condenser is shown in Fig. 1. It consists of a cylindrical cast iron body with a cone bottom section connecting with the tail-pipe. The vapors are led into the condenser through an elbow in the bottom portion of the condenser. Cooling water is admitted to the condenser near the top and is distributed in a cylindrical sheet by means of a serrated cone frustum. The falling sheet of water strikes a second conical distributor and then passes into the condensing space where the condensation of the incoming vapors takes place.

It is in this space that the non-condensable gases are separated and then removed through the dry air pump suction pipe which is attached to the top of the condenser. In passing from the condensation chamber to the air pump suction pipe, it will be seen that these gases must pass through the sheet of cold incoming condensing water formed between the cone frustum and the lower cone, resulting in a thorough washing and cooling of the gases, thus reducing their volume and the subsequent load on the dry air pump, and at the same time washing out any corrosive gases which may cause trouble in the air pump valves and cylinder.

The salt extractor is also of circular cross section and built of cast iron. The salt box is provided with a perforated cast iron false bottom on which rests a brass or bronze screen which has perforations fine enough to retain the salt crystals. When the extractor is filled, as seen through the sight glasses, the valve connecting it with the evaporator bottom is closed and the adhering liquor is blown from the crystals by means of steam admitted into the top of the extractor, this liquor being returned to the evaporator. The dry salt is then removed through a swinging door in the front of the salt extractor. On larger evaporators it is customary to provide two salt extractors so as to make the removal of the salt from the evaporator continuous instead of intermittent.

Chapter II.

The Manistee Evaporator.

(Courtesy of the Manistee Iron Works.)

The Manistee Evaporator was originally designed by Mr. George Ray of Manistee, Michigan, and therefore takes its name from the place of its birth. It was primarily intended for the evaporation of sodium chloride brine in the production of common salt, and has therefore been especially studied for adaptation to that particular problem, although it has been used in a few other industries.

In principle, this apparatus, as will be readily understood by referring to Figs. 1 and 2, is simply a modified standard effect in which mechanical circulation has been provided. Instead of having single or even double vapor pipes, it has quadruple inlets for the admission of steam in each calandria. This has been discussed in the chapter on Types of Steam Belts, to which the reader is referred. The matter of mechanical circulation has also been fully discussed in the chapter on that subject in the first section of this book.

There are some very interesting features in the design, not the least of which is the enormous size to which it has been built—as large as 30 feet in diameter. It is said, however, that this is not the largest in the world, Mirrilees Watson, of Glasgow, having built a multiple effect 36 feet in diameter. This size feature presents construction, erection and transportation problems quite out of the ordinary. Very considerable expansion and contraction strains are unavoidable, and have to be taken care of. The tube sheets are made in six to twelve sections with expansion joints between each, as it will be readily understood that upon admitting steam into the apparatus, it will take quite a while until all parts are heated equally, no matter how careful the operation, and unless this feature is taken care of, there will be breaks in the parts submitted to the maximum stresses.

For such enormous proportions, the castings must be made very heavy, and sectionalized into many pieces for shipment, requiring a great deal of fitting and matching. The weights to be supported are tremendous, not only as regards the pans, but also their brine content. It is therefore customary to provide independent columns for each body, sometimes as many as eight or twelve, all properly cross-braced and hog-tied.

As to the proportions, the tubes are usually five feet long, either 2 inches or 2½ inches diameter expanded directly into the tube sheets which extend through the flanges. The downtakes are about one-half the pan diameter.

The circulator is about 75 per cent. of the downtake diameter, and sets at a point slightly below the lower tube sheet. This circulator is merely a towing propeller, the blades being bolted to a block attached to the shaft,

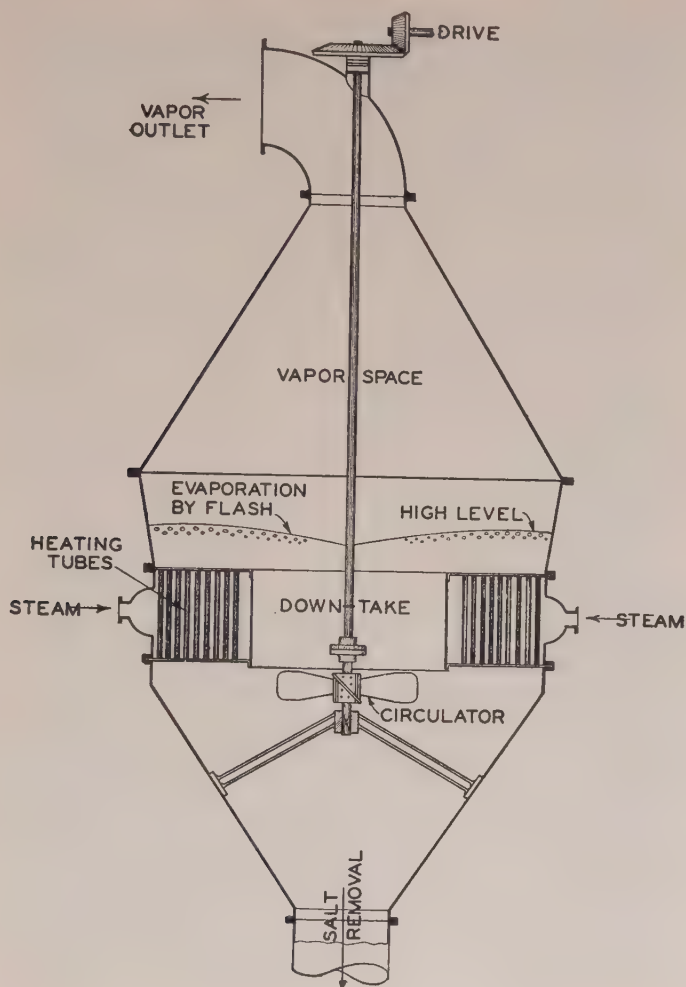


FIG. 1.

resting in a bearing below provided with wearing rings of the renewable type. The thrust and weight of this stirring mechanism is supported from the top. The drive above is furnished by a bevel gear transmission, with a speed reducer on the pinion shaft with a motor drive. The speed of the circulator is about 40 revolutions per minute. The direction of circulation

is up the tubes, and down the central tube, as it is with natural circulation. In actual operation, reversal of this direction does not seem to make much difference. It would seem that with reversed circulation, there would be

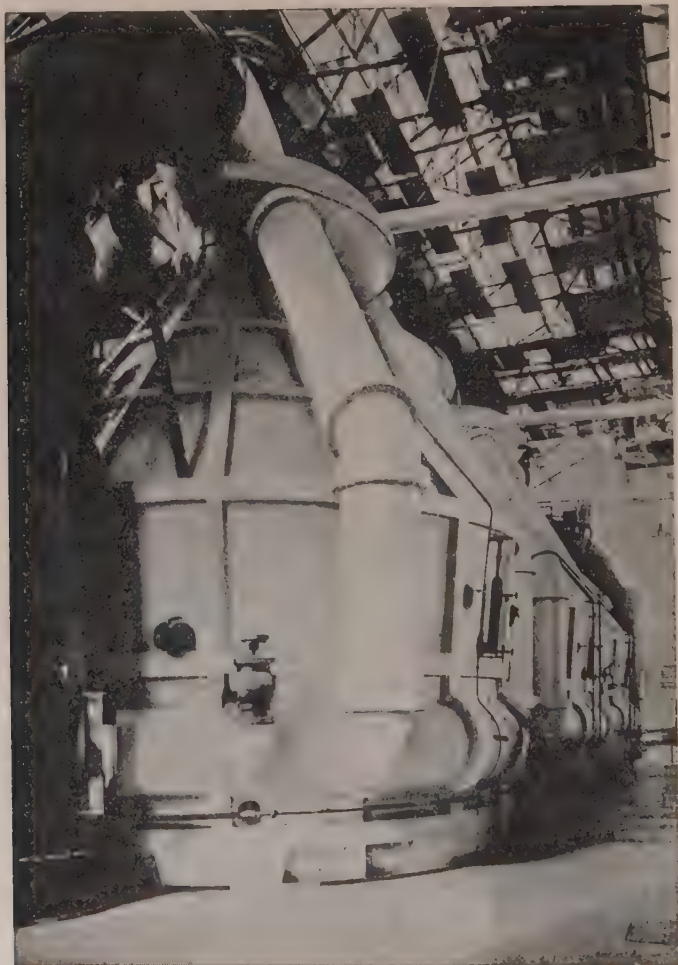


FIG. 2.—Manistee Evaporator.

less disturbance in the bottom, and therefore uninterrupted settling of salt.

While on this subject, it is characteristic of the Manistee Evaporator to make fairly coarse grain, which is a very desirable feature, and it may well be that the thrust of the propeller below keeps the small grain in

suspension, allowing them to grow, only the larger ones being able to settle against the currents from the back wash of the propeller. If such is the case, then it would be an error to reverse the circulation.

The lower cone is very steep and deep, permitting an uninterrupted settling of the salt crystals, without sticking to the sides. This salt is removed from the bottom in some approved manner, either by the use of a salt boot and elevator, or by means of a pump and outside settler.

The liquor belt immediately above the steam belt is flared out to a diameter of perhaps two feet more than the calandria diameter in a height of about five to six feet. The object of this arrangement is said to be the provision of a sort of shelf for the support of salt which accumulates on the sides of the liquor belt at the boiling level in the form of a cake. With a perfectly cylindrical shape here, there is danger of cracking this cake, and dropping it on the tube sheet, from which it is swept into the downtake, and in the bottom, where it might clog the salt outlet, necessitating a shut down with costly interruption.

From the top of the flare above, the dome extends upward in the form of a steep cone to the vapor outlet. The height above the boiling brine is therefore very high, and with the low rates of evaporation, and the very large vapor liberating area, it has been found unnecessary to provide against entrainment, which does not occur unless the solution being evaporated is of a foamy character, where, due to the nature of the evaporation (by flashing), it will cause serious difficulty, the froth reaching the top, and going out with vapor either into the succeeding steam belt or into the condenser with consequent loss in either case.

When used in the manufacture of salt from saturated brine, which is usually the case, each body is fed independently, the brine entering either above or below the calandria. When the various bodies are fed in series, the liquor enters on one side of each body and leaves on the other either from above or below the calandria, which probably does not make much difference in view of the mechanical circulation.

The Manistee evaporator has found many applications in the salt industry, and for the evaporation of various crystallizing solutions. It is found in very large sizes in many of our large industrial plants, where it has met with good success and is well thought of.

Chapter 12.

Distilled Water Evaporators.

With the growth and development of power plants to the present large capacities and high efficiencies, and also with the increased cost of fuel, engineers have found it necessary to study the operation of their boilers with great care. One of the results of this research work has been to show

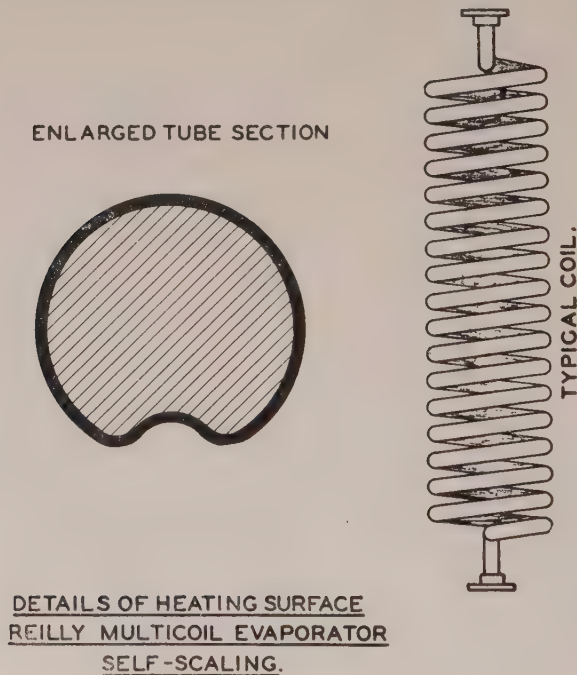
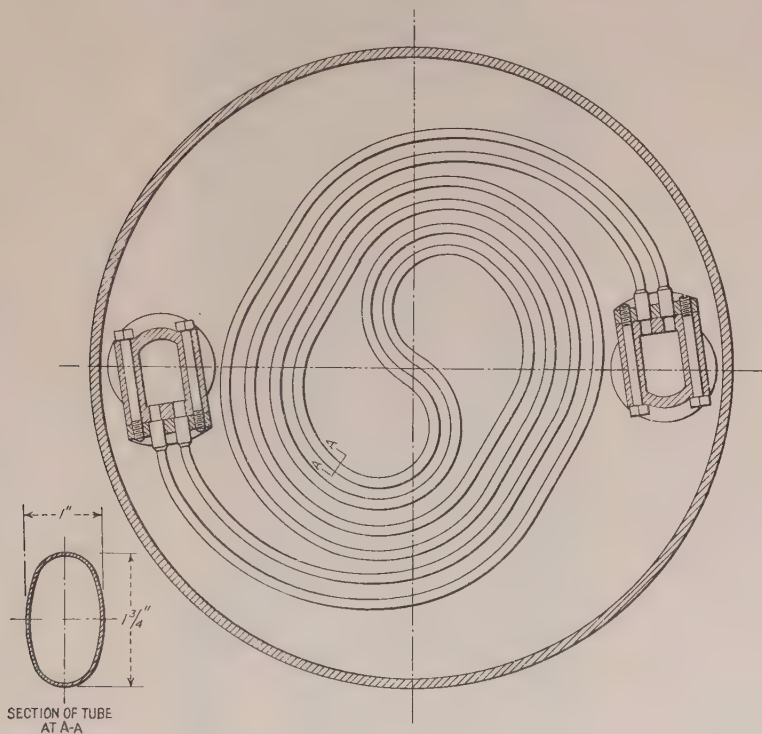


FIG. 1.—Cross Section Showing Shape of Coils.

that even slight accumulations of scale on the inside of steam boilers bring about a material reduction in their efficiency. Most of the water fed into these boilers consists of steam condensed in the surface condensers, but the inevitable losses have to be made up from some other source. A demand has therefore arisen for apparatus suitable for the production of distilled

water from raw water to supply this deficiency, and thus avoid introducing into the steam generating plant, boiler feed water containing scale or other impurities which might interfere with good operation.

The above is the greatest field for apparatus of this kind, but not the only one. In a good many tropical countries, the drinking water available



Courtesy Wheeler Condenser & Engineering Co.

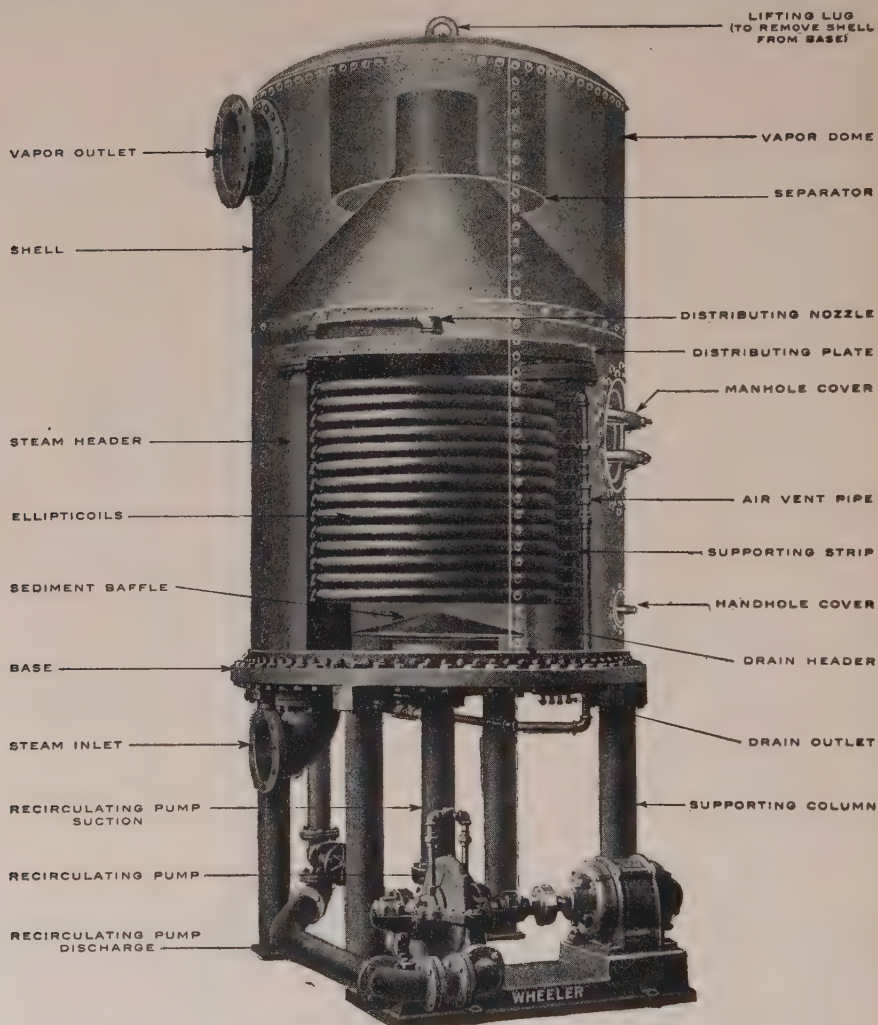
FIG. 2.

is very often unsuitable, and very often this is overcome by supplying distilled water properly aerated, which has proved to be very good, even though it does not contain any lime salts usually considered necessary in good drinking water.

In marine work, it has always been common practice to use distilled water for make-up purposes in the boilers to avoid the introduction of salt water into the system, and incidentally, the accompanying scale. The adaptation of this system to stationary plants has therefore had a well-established precedent.

The big problem in this work, and practically the only one lies in the fact that if the prevailing feed water is of such a character that it will

scale up the tubes of the steam boilers, it will also scale up the evaporators in the same way. The apparatus to do this work must therefore be designed and operated with this fact in mind.



Courtesy Wheeler Condenser & Engineering Co.

FIG. 3.

As has been stated in the description of the Lillie Evaporator, it has been found that by reversing the sequence of multiple effect evaporators both as to steam and as to direction of feed flow, making the first body

become the last, and the last, the first, most of the trouble from fouling is eliminated due to the cracking and falling of the scale from the tubes caused by changes of temperature, concentration, and deformation of the shape of the surface as compared with the shape of the scale clinging to it. Evaporators of this type are in common use in many large plants both stationary and marine.

Another method of avoiding the accumulation of scale on heating surfaces of apparatus of this sort and one in quite general use is known distortion or deformation of the shape of the surfaces.

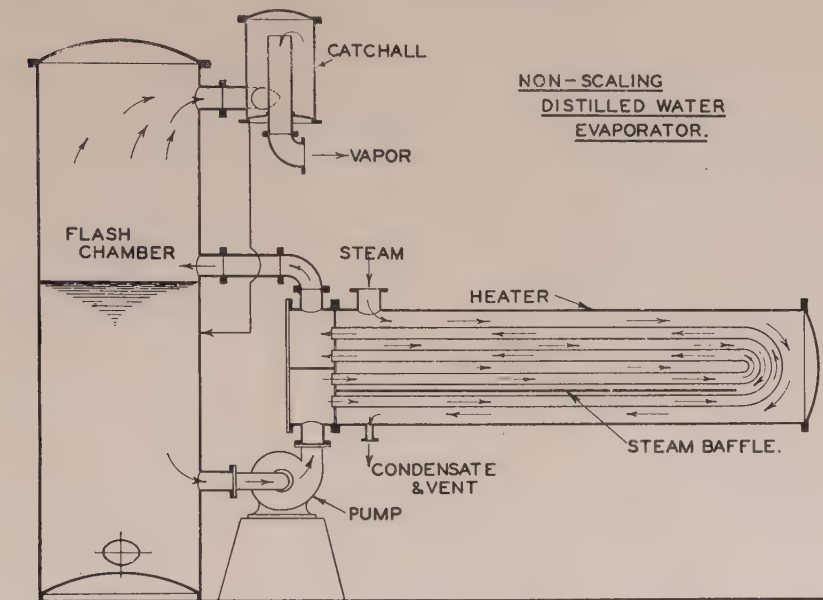


FIG. 4.

There are several designs in use to-day which successfully accomplish this task. In some, the surface is made up of a number of solenoidal coils made of copper tubes, steam being on the inside, and water to be evaporated on the outside. The tubes from which these coils are made are not perfectly circular in section, but of the shape shown in Fig. 1. It will be readily understood that under variations of temperature and pressure, there will be a material change of the curvature of the cross section, with sufficient distortion to cause the falling off of the attached scale.

Another design belonging to this same family consists of a number of flat spiral coils set one above the other, and designed in such a way that variations of pressure will cause corresponding changes in the shape of the spiral much on the order of the action of the elliptical tube on a Bourdon pressure gauge. Such a coil is shown in Figs. 2 and 3. In actual practice,

the design has proven successful in dropping off the scale as fast as it is formed, giving fairly good heat transmission.

There is still another design employing entirely different principles to accomplish the same result. This is a European invention based upon a modification of mechanical circulation in evaporators. It is known that when the velocity of circulation of liquid past the heating surface exceeds a certain minimum, no scale is deposited. This has been brought out in the chapter on Multicurrent Solution Heaters. The evaporator in question consists of a tubular heater in which the temperature of the water to be evaporated is raised above the boiling point under the conditions existing, after which this overheated water is discharged into flash or evaporating chamber, giving up the absorbed heat in excess of the temperature corresponding to the pressure existing at that point with consequent evaporation. From the bottom of this chamber, this water is again pumped through the heater by means of an appropriate pump for another circuit. The critical determination is the speed of this circulation through the tubes, which is believed to be in the neighborhood of 15 to 20 feet per second. Fig. 4 shows a diagram of the equipment.

It is probable that these methods of operation for scale prevention in evaporators will see considerable developments in the future, and find new and useful applications in other industries, where difficulties of similar nature are encountered.

Chapter 13.

The Zarembo Patent Evaporator

(Courtesy of the Zarembo Co.)

HORIZONTAL TUBE, ROUND BODY

In 1906 and 1907 Zarembo brought out the horizontal tube round body evaporator which was designed to combine the great operating advantages of the horizontal rectangular type with the structural advantages of the vertical tube round body type and included some new elements not found in either form of apparatus. The Zarembo Evaporator stands out both from a study of its design and operation and also from a standpoint of the results which it yields in operation and maintenance.

Operation. The vital element of any steam evaporator is the heating surface—its material, shape and position, and the conditions in which it works, that is, liquid circulation on one side—steam circulation on the other side. Other factors are important, but the heating surface is the vital factor.

Circulation. The most important requirement is vigorous and positive circulation. Rapid movement of the boiling liquid past the tubes means large heat transmission through the tube walls and correspondingly large evaporation per square foot of heating surface. In the ZPE the circulation is produced in the same manner as in the air-lift pump. The space around the tubes and between the partition plates at each side of the tube-nest is filled with a mixture of liquid and myriad bubbles of steam. The steam escapes at the surface while the remaining liquid rolls over into the down-takes at each side. Because of the excess weight of the two columns of liquid in the downtakes a constant pressure is produced underneath the tubes which forces the liquid past them rapidly and steadily.

To maintain a vigorous circulation it is necessary that there be provided a rapid and steady flow of steam through all the tubes and at the same time that there exist a complete contact between the flowing steam or vapor and the entire surface of the steam side of the tubes. Again, presence of air in the heating system of an evaporator affects seriously its capacity (same as in a steam radiator). The air dissolved in boiler feed water, plus leakage (if any) into the evaporator bodies is being entered continually into the evaporator steam chests and should be eliminated at once to secure high duty. Proper ventilation is impossible where steam is permitted to wander at will among a forest of tubes with no means provided for cornering and removing the insulating air. In the ZPE the steam drives its associated

air down an alley and corners it in the front steam chest where the air is collected by a special device and removed from the heating system. The steady rush of steam through the straight small tubes of the ZPE provides

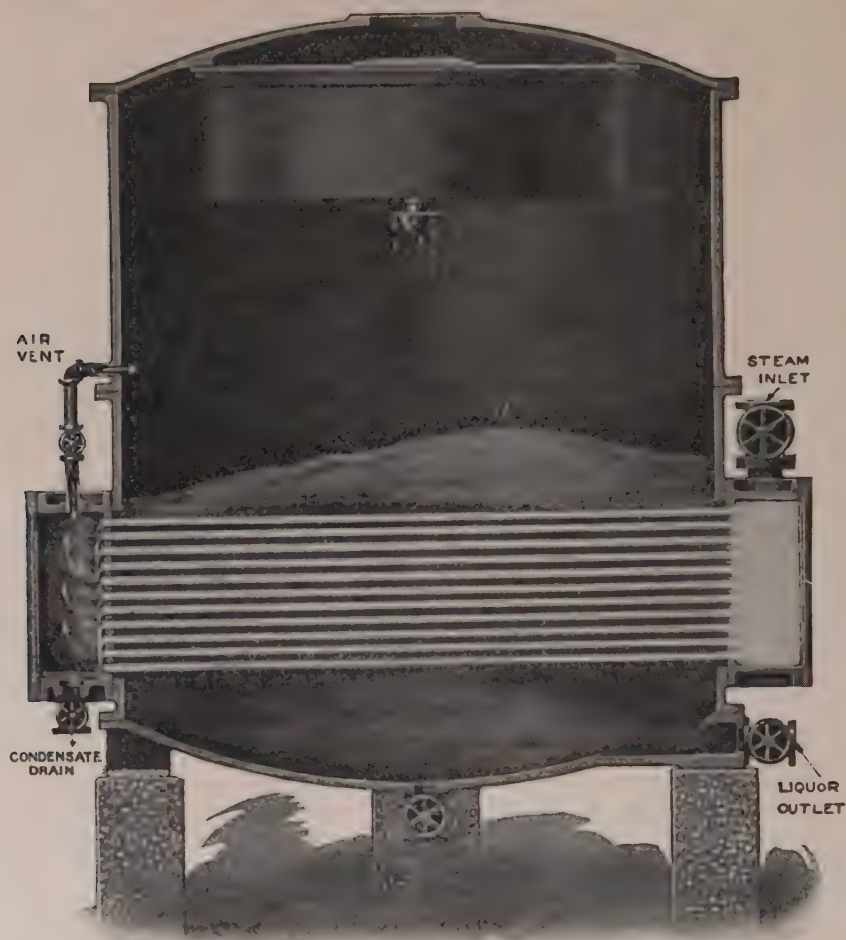


FIG. 1.—Typical Section Horizontal Tube Zarembo Evaporator.

these conditions, purging the tubes completely of the air and water of condensation. The evacuation of the condensation is made positive and complete by the downward pitch of the tubes.

The Fastening. The tubes in the Zarembo Patent Evaporator are secured in place by the use of elastic gaskets at each end of each tube, compressed by bolted packing plates. All tube holes are drilled to template



FIG. 2. Air Lift Principle.

and counterbored, hence perfect spacing of tubes and perfect fitting of gaskets is secured. The tubes are made tight by screwing down the stud-nuts upon the packing plates. The pressure causes the packing rings to compress the rubber gaskets which are then entirely surrounded by metal. No opening is left through which the rubber can ooze away. The joint is the equivalent of a pump rod or valve stem packing box.

Tubes can be quickly removed and readily replaced by unskilled labor. Old gaskets are easily removed without the use of a special tool; future increase in capacity can be provided for—requiring only the insertion of additional tubes.

Periodic cleaning is essential in all cases. The tubes are cleaned by boiling with diluted acid or alkali, or one after the other, followed by boiling in clear water. The piping of the ZPE can be so arranged that all bodies of a multiple effect can be cleaned, one at a time, without stopping operations.

Control. The ZPE, being of the submerged tube type and carrying a relatively large mass of liquid in each body, is subject to easy and accurate control. Good illumination of body makes possible a full knowledge of what is taking place inside, permitting elimination of irregularities in boiling due to fluctuations in steam supply, improper position of liquid level or to scaling of heating surface.

Due to the volume of liquid present, a uniform density of liquid discharge is secured in spite of variations in rate of feed, steam pressure or vacuum. Where operators are careless an automatic liquid level controller can be depended upon to keep the liquid at the chosen height. Illumination of body is secured by use of internal incandescent bulbs, placed immediately below the top head, each shielded by a heavy glass globe surrounded by a wire basket.

Losses. It behooves the prospective purchaser of an evaporator to investigate carefully the provisions made for the prevention of entrainment loss, in view of the frequent claim of manufacturers that they cause no such loss. Speaking generally, it is a difficult matter to stop entrainment after the vapor has once entered the vapor pipe and has thereby greatly accelerated its speed. Any separation of liquid particles from the vapors should be performed while those vapors are still moving at comparatively slow speed, that is, in the evaporator body and before entering the vapor pipe. One method for doing this is to provide vapor space over the boiling liquid, thus affording the force of gravitation sufficient time in which to pull back such liquid particles as may have been projected upwards. This action is facilitated by providing a proper vapor-liberating area for the separation of the vapor from the boiling liquid. In comparing the liberating area of various evaporators attention must be paid not only to the cross section of the body, but also to the rate at which the boiling surface renews itself—that is, the speed of liquid circulation.

In the ZPE the vapor space (the space above the tube-nest) is given ample height and is so arranged that its entire volume is used effectively to prevent entrainment loss. As an additional safeguard against carelessness in operation, each body is fitted with an internal separator which

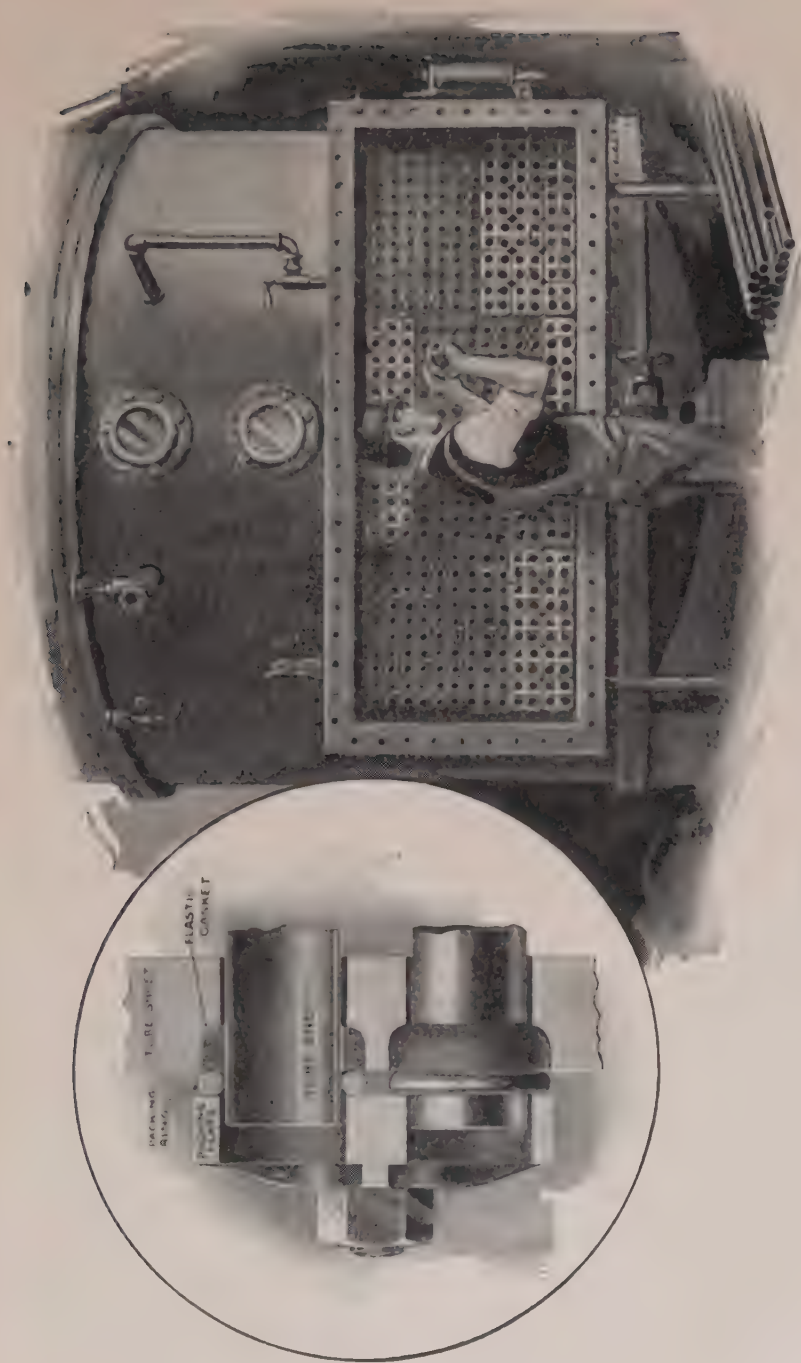


FIG. 3.—Method of Fastening Tubes in Zarembo Evaporators.

returns to the boiling liquid particles that may have been projected upward into the vapor space. This separator is much more effective than the old style catchalls and baffle plates.

The liquor loss by creepage (upward movement of liquid on internal surface of body by dragging effect of high speed vapor) is frequently unrecognized in the operation of old style bodies. Losses of this character are impossible where internal separators, either of the bowl type or centrifugal type are used.

Centrifugal Separator. Baffle plate arrangements and various forms of separators have been used since the first evaporator was built but with unsatisfactory results. The Hughes Centrifugal Separator and Foam Arrester is of unusual importance to those who experience losses from entrainment or foaming. In this device the escaping vapor passes through the long spiral passageway at high velocity developing a centrifugal force many times that of the pull of gravity, and throwing entrained liquid or masses of foam to the outer edge of the vapor path where it comes in contact with the wall of the spiral passageway, a specially perforated plate having innumerable small fins which catch and retain the liquid particles and in the case of foam, exert a slicing action which reduces it to a liquid by breaking up the foam bubbles. The liquid trickles down the perforated wall of the separator to the bottom (being out of contact with the high-speed vapor) flows to the sealed drain under the center of the separator and is returned to the evaporator. When the vapor reaches the end of the long spiral path all foam has been destroyed and all entrained liquid has been recovered from the vapor. The efficiency of the separating action is due to the high velocity, the length of the spiral passageway, and the peculiar construction of the perforated wall.

The centrifugal separator has been well tried out in actual practice in evaporators from the smallest to the largest size. It can be attached to existing installation of any make. It is built for external connection to old evaporators where difficulty would be experienced in mounting it internally. Its use is recommended for any liquid subject to loss from entrainment or foaming such as sugar, glycerine, glue, soda and sulphate black liquor, caustic soda or soda recovery.

Structure and Maintenance. **Strength and Safety.** A worth-while evaporator body carrying vacuum must keep the atmosphere in its place—outside—each day and every day for a long period of years. This condition requires that the body joints be made tight in erection and easily kept tight in operation. Zarembo bodies have but three joints and those of the simplest type—flanged joints with broad, finished faces and drilled bolt holes, 4 inches apart. These joints are made up with asbestos tape carefully lapped. No tee joints or corner joints are used, thereby eliminating the leakage that is otherwise unavoidable. Result: A vacuum that stands for many hours after the pumps are stopped.

Zarembo bodies are made cylindrical for the same reason that boilers are so made—no body of this make has ever “blown up” or collapsed. They are built to be safe companions with which to work.

Each body is a self-contained structural unit, so that should settling

of foundations occur, the resulting damage is confined to the piping, the bodies being in no wise injured.

Heating Surface. The exceedingly important factor of tube sheet design—the action of large, flat, perforated plates under varying degrees of pressure—has been carefully and thoroughly investigated by Witte. The results of this extended research are incorporated in the ZPE. All tube holes in tube sheets are drilled to template and carefully counterbored to provide perfect fitting gaskets; the tube rests used in the larger bodies, are also drilled to template. Hence, all tubes are uniformly spaced and accurately centered in the tube rests. The rubber gaskets, when compressed

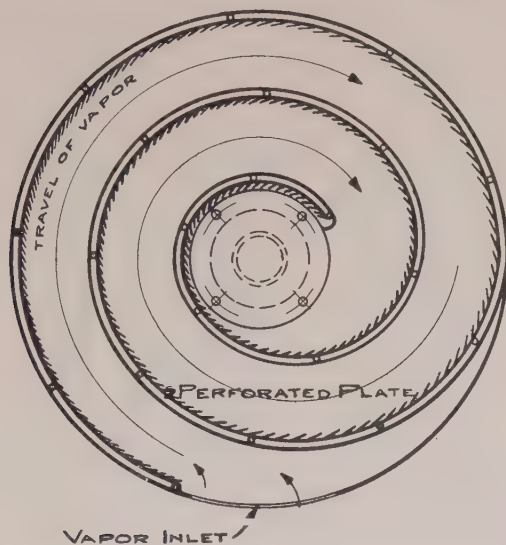


FIG. 4.—Diagram Showing Operation of Hughes Centrifugal Separator.

by packing rings placed underneath the packing plates are completely surrounded by metal, permitting no portion of the gasket to escape from the packing and thereby produce an imperfect joint.

During the long life of Zaremba bodies, it is sometimes necessary to replace tube sheets, because of accidents in operation or corrosion by acid liquor. Renewal is affected by removal of the steam chest only—no dismantling of body is necessary. In all bodies the steam chest covers are fitted with one or more manhole plates, thus affording easy access to interior of steam chest for inspection of tube joints and setting up of gaskets. One man can do the work, whereas, otherwise the use of several would be necessary.

Economic Factors. Space Required.—The use of small diameter ($1\frac{1}{4}$ inches) tubes gives large heating surface per cubic foot of volume, while the use of the Centrifugal Separator makes possible a substantial

reduction in height of vapor space. Because of these two considerations the ZPE occupies relatively small floor space and can be installed where the head room available is scant.

Heating Surface. The long tubes used can expand and contract without resistance, by movement through the packing rings, thereby avoiding the injury experienced by tubes held rigidly in position by being rolled into tube sheets. Zaremba tubes are accurately spaced (reducing resistance to liquor circulation and heavy liquor discharge) and kept straight by the use of tube rests accurately registered. Due to the inclined position of tubes all condensation is immediately discharged no water stands in the tubes (to cause corrosion) when work has been stopped. The tube gaskets, being surrounded completely by metal cannot ooze away when the packing plates are screwed down. All tube joints can be made tight and kept tight, hence no leakage, either of steam from steam chest into liquid space, during operation, or of liquid into steam chest during idleness of apparatus.

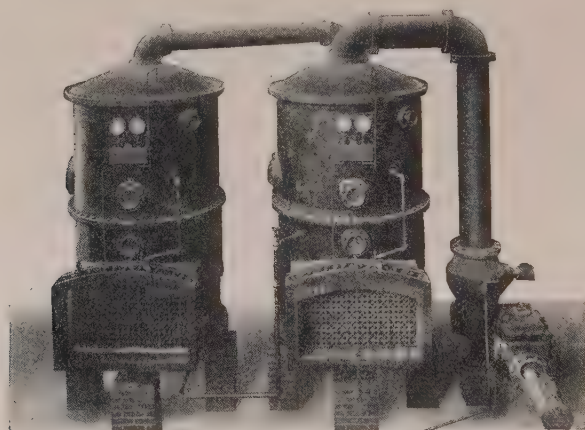


FIG. 5.—Zaremba Patent Evaporator. Lined Double Effect for Garbage Tank-water.

Tube Replacement. All tube replacement can be performed during nights or Sundays—hence the operating period of the ZPE is not interfered with. To remove old tubes and replace with new ones requires the removal of steam chest covers and packing plates only.

Covering. To prevent loss of heat by radiation an insulating covering must be applied overall. In the case of a ZPE cast-iron box body 95 per cent. of its surface can be covered without difficulty and still leave all bolts accessible.

Linings. Where liquids are concentrated in acid condition it is necessary that the apparatus be able to resist the attack of acid and that the liquid be not contaminated by the dissolved metal. To provide this resistance it was formerly necessary to construct the evaporator bodies of copper, lead or aluminum, calling for great additional expenditure as com-

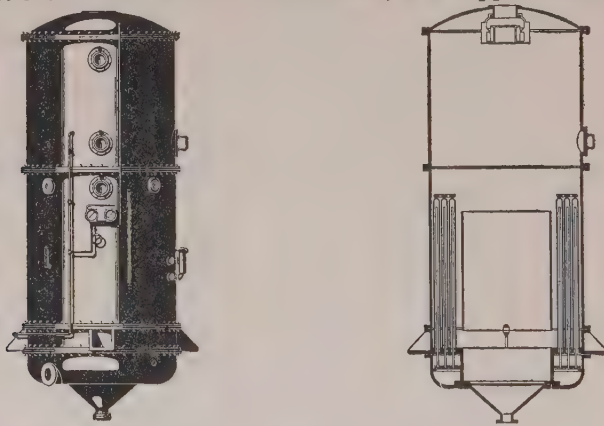
pared with the customary cast iron or steel construction. The Zaremba System of Acid-proof lining for evaporators makes possible the use of iron or steel evaporator bodies for concentrating acid liquids. It comprises acid-proof brick, or tile, set in and covered by a cement resistant to the acid present. The surfaces are so designed that the lining is given an arched construction at all points thereby making it impossible for accidental, unbalanced pressure on the lining to disturb it. During eleven years of experience with this system there has been not one case of rupture. These Acid Evaporators include all the features of design peculiar to the Zaremba Patent Evaporator and provide an apparatus of the utmost degree of simplicity both as regards installation and operation. The picture on page 468 shows a small Zaremba double effect for garbage tank water containing acetic and fatty acids. After four years of use it was dismantled for removal to another location, the bodies being broken apart at the middle joint for ease in handling. The lining retained its position perfectly, acting as though it were homogeneous with the cast iron belt.

General. The facts stated in this chapter cause Zaremba Co. to feel amply warranted in making the general claim for their horizontal tube round body evaporator—"built for hard work, long life and no worry to its owner." Much credit is due to Hughes, Mantius, Witte, Eggert and Pancoast for improvements in detail of body and auxiliaries and arrangement of installations. The Zaremba Evaporator has been patented 1914 and 1921, other patents pending. The Hughes Centrifugal Separator has been patented 1925.

Chapter 14.

High Concentrator.

For the concentration of solutions of extreme densities and high boiling points, there has been developed a type of apparatus as shown in Fig. 1. In this arrangement, the tubes are attached to the tube sheet at the lower part of the tubes only, the upper ends being closed and free for expansion and contraction. Below this main tube sheet, is a supplemental one having



Courtesy Buffalo Foundry & Machine Co.

FIG. 1.—High Concentrator.

smaller tubes corresponding to the main tubes, extending vertically and threading into these. Steam is admitted from the space below the supplemental tube sheet, enters through the small tubes, and is thus discharged into the upper ends of the main evaporating tubes. As condensation takes place, the flow is downwards in the annular space between the small and large tubes, the condensate and gases being removed from the space between the two tube sheets.

The tubes are placed in a circular band close to the outer periphery of the evaporator, and just inside of this annular bank is a cylindrical apron extending to within a short distance of the bottom, enclosing a space which serves as a downtake. Thus the path of circulation is well defined indeed. This design is very suggestive of the Sanborn Evaporator, except that the direction of steam flow through the tubes is reversed. The transmission of heat should be very good.

The apparatus is used successfully for the high concentration of caustic soda, calcium chloride, and similar materials, where evaporation is carried so far that solidification takes on moderate cooling. The intelligent use of this apparatus has made possible the elimination of the direct fired cast iron pot for work of this nature in many cases.

The design is arranged for operation with high pressure steam on one side, and vacuum on the other, thus giving a large temperature range, which is indispensable in work of this character.

Appendix.

In the following pages is given a series of tabulated data useful in working out certain evaporator problems. The figures have been made up for this work without carrying the extensions too far, as there is no necessity in this work, the readings having been taken from ordinary indicating instruments such as are used in practice. The information has been grouped for convenience. The sources of the data are various, but principally the Chemists' Year Book for 1925. The following is the list:

1. Abbreviated Steam Tables;
2. Various Metric Equivalents;
3. Conversion Chart for Fahrenheit, Centigrade, and Reaumur temperatures;
4. Specific Gravities corresponding to various density spindles;
5. Boiling Point Rise of various solutions at atmospheric pressures;
6. Viscosities of sugar solutions at various temperatures;
7. Solubility of air in water at various temperatures;
8. Solubilities of various compounds by groups with specific gravities of solutions and corresponding solid contents.

PROPERTIES OF SATURATED STEAM AT DIFFERENT VACUA AND GAUGE PRESSURES.

(Referred to 30-in. Barometer)

Vacuum	Temp. F.	Cu. ft. per Lb.	Latent Heat
28	101.15	339.6	1,035
27.5	108.70	275.2	1,030
27	115.06	231.9	1,027
26.5	120.55	200.2	1,024
26	125.38	176.7	1,021
25.5	129.75	158.1	1,019
25	133.77	143.0	1,017
24	140.64	119.0	1,013
23	146.78	104.5	1,009
22	152.16	92.3	1,006
21	157.00	82.6	1,003
20	161.42	74.8	1,001
19	165.42	68.5	998
18	169.14	63.1	996
17	172.63	58.6	994
16	175.93	54.5	992
15	179.03	51.2	990
14	181.92	49.0	989
13	184.68	45.5	987
12	187.31	43.2	985
11	189.83	41.1	984
10	192.23	39.21	983
9	194.52	37.4	981

APPENDIX

(Referred to 30-in. Barometer)

Vacuum	Temp. F.	Cu. ft. per Lb.	Latent Heat
8	196.73	35.8	980
7	198.87	34.3	978
6	200.94	33.0	977
5	202.92	31.8	976
4	204.85	30.6	975
3	206.71	29.5	974
2	208.52	28.6	973
1	210.28	27.7	972
0	212.0	26.8	970

Pressure (lbs.)	Temp. F.	Cu. ft. per Lb.	Latent Heat
1	215.3	25.2	968
2	218.5	23.8	965
3	221.5	22.5	964
4	224.4	21.4	962
5	227.2	20.4	961
6	229.8	19.5	959
7	232.4	18.6	957
8	234.8	17.8	955
9	237.1	17.1	954
10	239.4	16.5	953
11	241.6	15.85	951
12	243.7	15.35	950
13	245.8	14.85	948
14	247.8	14.35	947
15	249.8	13.8	945
16	251.6	13.45	944
17	253.4	13.1	943
18	255.2	12.7	942
19	257.0	12.35	940
20	258.8	12.0	939
21	260.4	11.7	938
22	262.0	11.4	937
23	263.7	11.1	936
24	265.2	10.85	935
25	266.6	10.6	934
30	274.0	9.46	929
35	280.7	8.58	924
40	286.9	7.81	920
45	292.4	7.20	915
50	297.7	6.68	911
55	302.7	6.23	908
60	307.3	5.83	904
65	311.8	5.49	901
70	316.0	5.19	897
75	320.0	4.91	894
80	323.9	4.67	891
85	327.6	4.44	888
90	331.2	4.24	885
95	334.6	4.06	883
100	337.9	3.89	880
105	341.1	3.74	878
110	344.2	3.59	875
115	347.2	3.46	873
120	350.0	3.34	870
125	352.9	3.26	868

BASIC METRIC EQUIVALENTS.

French	British and American	British and American	French
1 cm.	= 0.3937 in.	1 in.	= 2.54001 cm.
1 meter	= 39.37 in. = 3.28083 ft. = 1.0936 yd.	1 ft.	= 0.304801 m.
1 sq. cm.	= 0.15500 sq. in.	1 sq. in.	= 6.4516 sq. cm.
1 sq. m.	= 10.7639 sq. ft. = 1.1998 sq. yd.	1 sq. ft.	= 0.092903 sq. m.
1 cu. cm. = 1 milliter	= 0.061023 cu. in.	1 cu. in.	= 16.387 cu. cm.
1 cu. m.	= 35.3145 cu. ft. = 1.308 cu. yd.	1 cu. ft.	= 0.028317 cu. m.
1 liter = 1 cu. dm.	= 0.26417 gals. = 61.0234 cu. in. = 0.03531 cu. ft. = 2.202 lbs. water at 62° F.	1 gallon	= 3.7854 liters
1 hectoliter (or decistere)	= 26.417 gals. = 3.5314 cu. ft.	1 gal.	= 0.037854 hectoliters
1 kilogram	= 2.204622 lbs.	1 lb.	= 0.453592 kilograms
1 gram/cu. cm.	= 62.4283 lbs./cu. ft.	1 lb./cu. ft.	= 0.016018 grams/ cu. cm.
1 kilogram per sq. cm.	= 14.223 pounds per sq. in.		
1 pound per sq. in.	= 0.070307 kilograms per sq. cm.		

DEGREES			DEGREES			DEGREES			DEGREES		
C.	F.	R.	C.	F.	R.	C.	F.	R.	C.	F.	R.
0	32	0	10	50	8	20	68	16	30	86	24
	33			51			69			87	
1	34	1	11	52	9	21	70	17	31	88	25
	35			53			71			89	
2	36	2	12	54	10	22	72	18	32	90	26
	37			55			73			91	
3	38	3	13	56	11	23	74	19	33	92	27
	39			57			75			93	
4	40	4	14	58	12	24	76	20	34	94	28
	41			59			77			95	
5	42	5	15	60	13	25	78	21	35	96	29
	43			61			79			97	
6	44	6	16	62	14	26	80	22	36	98	30
	45			63			81			99	
7	46	7	17	64	15	27	82	23	37	100	31
	47			65			83			101	
8	48	8	18	66	16	28	84	24	38	102	32
	49			67			85			103	

Chart for Temperature Conversions.

DEGREES			DEGREES			DEGREES			DEGREES		
C.	F.	R.	C.	F.	R.	C.	F.	R.	C.	F.	R.
40	104	32	50	122	40	60	140	40	70	158	56
	105			123			141			159	
41	106	33	51	124	41	61	142	40	71	160	57
	107			125			143			161	
42	108		52	126		62	144		72	162	
	109	34		127	42		145	50		163	56
43			53			63			73		
	110			128			146			164	
44	111	35	54	129	43	64	147	51	74	165	59
	112			130			148			166	
45	113	36	55	131	44	65	149	52	75	167	60
	114			132			150			168	
46	115		56	133		66	151		76	169	
	116	37		134	45		152	53		170	61
47			57			67			77		
	117			135			153			171	
		38			46		154	54			62
48	118		58	136		68			78	172	
	119			137			155			173	
49		39	59	138	47	69	156	55	79	174	63
	120						157			175	
	121			139							

Chart for Temperature Conversions.

DEGREES			DEGREES			DEGREES			DEGREES		
C.	F.	R.	C.	F.	R.	C.	F.	R.	C.	F.	R.
80	176	64	90	194	72	100	212	80	110	230	80
	177			195			213			231	
81	178	65	91	196	73	101	214	81	111	232	89
	179			197			215			233	
82	180		92	198		102	216		112	234	
	181	66		199	74		217	82		235	90
83	182		93	200		103	218		113	236	
	183	67		201	75		219	83		237	91
84	184		94	202		104	220		114	238	
	185	68	95	203	76	105	221	84	115	239	92
	186			204			222			240	
86	187	69	96	205	77	106	223	85	116	241	93
	188			206			224			242	
87	189		97	207		107	225		117	243	
	190	70		208	78		226	86		244	94
88	191		98	209		108	227		118	245	
	192	71	99	210	79	109	228	87		246	95
89	193			211			229		119	247	

Chart for Temperature Conversions.

CORRESPONDING SPECIFIC GRAVITY FOR SPINDLE READINGS OF
VARIOUS HYDROMETERS.

Spindle Reading	Twaddell °	Barkometer	Salinometer ^a	Baumé ^a (Amer.)	Brix ^b
1	1.005	1.001	1.002	1.007	1.004
2	1.010	1.002	1.003	1.014	1.008
3	1.015	1.003	1.005	1.021	1.012
4	1.020	1.004	1.007	1.028	1.016
5	1.025	1.005	1.009	1.036	1.020
6	1.030	1.006	1.010	1.043	1.024
7	1.035	1.007	1.012	1.051	1.028
8	1.040	1.008	1.014	1.058	1.032
9	1.045	1.009	1.016	1.066	1.036
10	1.050	1.010	1.017	1.074	1.040
11	1.055	1.011	1.019	1.082	1.044
12	1.060	1.012	1.021	1.090	1.049
13	1.065	1.013	1.023	1.098	1.053
14	1.070	1.014	1.025	1.107	1.057
15	1.075	1.015	1.026	1.115	1.061
16	1.080	1.016	1.028	1.124	1.066
17	1.085	1.017	1.030	1.133	1.070
18	1.090	1.018	1.032	1.142	1.074
19	1.095	1.019	1.034	1.151	1.079
20	1.100	1.020	1.035	1.160	1.083
21	1.105	1.021	1.037	1.169	1.088
22	1.110	1.022	1.039	1.179	1.092
23	1.115	1.023	1.041	1.188	1.097
24	1.120	1.024	1.043	1.198	1.101
25	1.125	1.025	1.045	1.208	1.106
26	1.130	1.026	1.046	1.218	1.111
27	1.135	1.027	1.048	1.229	1.115
28	1.140	1.028	1.050	1.239	1.120
29	1.145	1.029	1.052	1.250	1.125
30	1.150	1.030	1.054	1.261	1.130
31	1.155	1.031	1.056	1.272	1.134
32	1.160	1.032	1.058	1.283	1.139
33	1.165	1.033	1.059	1.295	1.144
34	1.170	1.034	1.061	1.306	1.149
35	1.175	1.035	1.063	1.318	1.154
36	1.180	1.036	1.065	1.330	1.159
37	1.185	1.037	1.067	1.343	1.164
38	1.190	1.038	1.069	1.355	1.169
39	1.195	1.039	1.071	1.368	1.174
40	1.200	1.040	1.073	1.381	1.179
41	1.205	1.041	1.075	1.394	1.185
42	1.210	1.042	1.077	1.408	1.190
43	1.215	1.043	1.079	1.422	1.195
44	1.220	1.044	1.081	1.436	1.200
45	1.225	1.045	1.083	1.450	1.206
46	1.230	1.046	1.085	1.465	1.211

^a Liquids heavier than water @ 60° F. Deg. Baumé = $145 - \frac{145}{\text{Sp. Gr.}}$

^b Mategezek and Scheibler. Sp. Gr. of solution at 17.5° C. where water at 17.5° C. = 1.

^c Castel-Evans, Physico-Chemical Tables. Sp. Gr. 1 + $\frac{5^\circ (\text{Tw})}{1000}$

^d Engelhardt, U. S. Mineral Resources.

CORRESPONDING SPECIFIC GRAVITY FOR SPINDLE READINGS OF
VARIOUS HYDROMETERS (*Continued*).

Spindle Reading	Twaddell °	Barkometer	Salinometer ^d	Baumé ^a (Amer.)	Brix ^b
47	1.235	1.047	1.087	1.480	1.216
48	1.240	1.048	1.089	1.495	1.222
49	1.245	1.049	1.091	1.510	1.227
50	1.250	1.050	1.093	1.526	1.233
51	1.255		1.095	1.543	1.238
52	1.260		1.097	1.559	1.244
53	1.265		1.100	1.576	1.250
54	1.270		1.102	1.593	1.255
55	1.275		1.104	1.611	1.261
56	1.280		1.106	1.629	1.267
57	1.285		1.108	1.648	1.272
58	1.290		1.110	1.667	1.278
59	1.295		1.112	1.686	1.284
60	1.300		1.114	1.706	1.290
61	1.305		1.116	1.726	1.296
62	1.310		1.118	1.747	1.302
63	1.315		1.121	1.768	1.308
64	1.320		1.123	1.790	1.314
65	1.325		1.125	1.812	1.320
66	1.330		1.127	1.835	1.326
67	1.335		1.129	1.859	1.332
68	1.340		1.131	1.883	1.338
69	1.345		1.133	1.908	1.345
70	1.350		1.136	1.933	1.351
71	1.355		1.138	1.959	1.357
72	1.360		1.140	1.986	1.364
73	1.365		1.142	2.014	1.370
74	1.370		1.144	2.042	1.376
75	1.375		1.147	2.071	1.383
76	1.380		1.149	2.101	1.389
77	1.385		1.151	2.132	1.396
78	1.390		1.154	2.164	1.402
79	1.395		1.156	2.197	1.409
80	1.400		1.158		1.416
81	1.405		1.160		1.422
82	1.410		1.163		1.429
83	1.415		1.165		1.436
84	1.420		1.167		1.443
85	1.425		1.170		1.450
86	1.430		1.172		1.457
87	1.435		1.175		1.464
88	1.440		1.177		1.471
89	1.445		1.179		1.478
90	1.450		1.182		1.485
91	1.455		1.184		1.492
92	1.460		1.186		1.499
93	1.465		1.189		1.506
94	1.470		1.191		1.513
95	1.475		1.194		1.521
96	1.480		1.196		1.527
97	1.485		1.198		1.534
98	1.490		1.201		1.542
99	1.495		1.203		1.549
100	1.500		1.205		1.556

BOILING TEMPERATURES OF SOME SOLUTIONS AT ATMOSPHERIC PRESSURE.^a

Substance	Per Cent. Anhydrous Solid 10%	20%	30%	40%	50%	60%	70%	80%	100%
Ammonium chloride (NH_4Cl)	215.1°	219.3°	225.7°	232.7°	46.9% 239.0°				
Ammonium nitrate (NH_4NO_3)	214.0°	216.5°	219.5°	223.5°	228.9°	243.4° 53.6%	256.0°	100% 464.0°	
Ammonium sulphate ($(\text{NH}_4)_2\text{SO}_4$)	213.3°	214.9°	217.3°	220.4° 35.5%	224.8°	226.8°			
Barium chloride (BaCl_2)	213.6°	215.5°	218.1°	220.1°					
Calcium chloride (CaCl_2)	215.4°	221.0°	230.5°	246.7°	266.0°	290.8°	324.5°	75.8% 356.0°	
Calcium nitrate ($\text{Ca}(\text{NO}_3)_2$)	214.0°	216.5°	219.7°	224.1°	230.5°	242.6°	264.7°	297.5°	82.0% 306.0°
Citric acid ($\text{C}_6\text{H}_8\text{O}_7$)	212.7°	213.6°	215.0°	216.9°	219.5°	223.5°	230.9°	244.0°	
Copper sulphate (CuSO_4)	212.2°	212.9°	214.3°	217.1°	219.6°	45.0% 215.2°			
Lead acetate ($\text{Pb}(\text{C}_2\text{H}_3\text{O}_2)_2$)	212.3°	212.6°	213.0°	213.5°	214.3°	215.2°	216.8°	219.0°	271.0°
Lithium chloride (LiCl)	218.0°	229.1°	252.0°	277.0° 38.5%	306.0°	334.0°			
Magnesium chloride (MgCl_2)	216.4°	224.2°	239.9°	266.0°					
Magnesium sulphate (MgSO_4)	213.1°	214.9°	218.1°	224.2°	42.7% 226.4°				
Oxalic acid ($(\text{COOH})_2$)	213.5°	215.6°	218.1°	221.9°					
Potassium acetate ($\text{KC}_2\text{H}_3\text{O}_2$)	215.3°	219.3°	224.2°	230.5°	240.5°	252.0°	271.0°	298.0°	86.3% 322.0°
Potassium carbonate (K_2CO_3)	213.6°	216.0°	219.7°	226.6°	237.7°	263.0°	272.0°		
Potassium chlorate (KClO_3)	213.5°	215.3°	217.3°	219.6°	41.3% 219.9°				
Potassium chloride (KCl)	214.2°	217.8°	223.0°	36.4% 227.3°					
Potassium hydrate (KOH)	216.3°	223.8°	236.3°	257.5°	292.0°	352.0°	441.0°	554.0°	
Potassium iodide (KI)	213.3°	214.9°	217.1°	220.1°	224.6°	231.8°	244.3°		
Potassium nitrate (KNO_3)	213.3°	214.9°	216.9°	219.4°	222.7°	227.1°	232.7°	79.5% 239.0°	

Substance	Per Cent. Anhydrous Solid									
	10%	20%	30%	40%	50%	60%	70%	80%	100%	
Potassium sodium tartrate ($\text{KNaC}_4\text{H}_4\text{O}_6$)	213.2°	214.7°	216.5° 23.1%	219.1°	222.6°	227.8°	235.2°	248.5°	329.0°	
Potassium sulphate (K_2SO_4)	213.4°	215.0°	215.6°							
Potassium tartrate ($\text{K}_2\text{C}_4\text{H}_4\text{O}_6$)	213.2°	214.7°	216.5°	219.1°	222.8°	228.5°	237.7° 67.5%			
Sodium acetate ($\text{NaC}_2\text{H}_3\text{O}_2$)	214.3°	217.8°	222.8°	230.4°	239.0° 47.6%	249.0°	257.0°			
Sodium borate ($\text{Na}_2\text{B}_4\text{O}_7$)	213.2°	214.5°	216.1°	217.8°	219.2°					
Sodium carbonate (Na_2CO_3)	213.9°	216.3°	219.5° 20.0%	221.0° 33.9%	223.0°					
Sodium chloride (NaCl)	215.3°	220.8°	227.8°	239.0°	241.7°					
Sodium hydrate (NaOH)	217.1°	226.4°	241.7°	263.0°	288.0°	319.0° 52.5%	358.0°	405.0°	597.0°	
Sodium hydrogen phosphate (Na_2HPO_4)	213.1°	214.6°	216.5°	219.0°	222.6°	223.7°	248.0°			
Sodium nitrate (NaNO_3)	214.2°	216.7°	220.1°	224.4° 31.7%	230.0°	239.0°	248.0°			
Sodium sulphate (Na_2SO_4)	213.0°	214.6°	217.1°	217.8°						
Sodium tartrate ($\text{Na}_2\text{C}_4\text{H}_4\text{O}_6$)	213.1°	214.6°	216.4°	219.0°	222.5°	59.3% 227.1°		75.0% 257.0°		
Sodium thiosulphate ($\text{Na}_2\text{S}_2\text{O}_3$)	213.4°	215.3°	217.9°	222.8°	229.4°	238.0°	250.0°	257.0°		
Strontium chloride (SrCl_2)	213.8°	216.5°	220.8°	228.7°	239.0°					
Strontium nitrate ($\text{Sr}(\text{NO}_3)_2$)	212.8°	213.9°	215.4°	217.7°	221.4°	53.7% 223.3°	221.6°	230.5°		
Sucrose ($\text{C}_{12}\text{H}_{22}\text{O}_{11}$)	212.2°	212.5°	213.1°	214.0°	215.4°	217.6°	221.6°	230.5°		
Tartaric acid ($\text{H}_2\text{C}_4\text{H}_4\text{O}_6$)	213.2°	214.6°	216.5°	218.9°	222.4°	227.3°	237.0°	250.0°	338.0°	
Zinc sulphate (ZnSO_4)	212.7°	213.8°	215.3°	218.1°	221.0°					

* Determinations by C. S. Robinson.

VISCOSITIES IN CENTIPOISES OF SUGAR SOLUTIONS.

Temp. C.	Grams. Sucrose in 100 Grams Solution.			
	0	20	40	80
0°	1.7920	3.804	14.77	238.0
10°	1.3077	2.652	9.7494	109.8
20°	1.0050	1.960	6.200	56.5
30°	.8007	1.504	4.382	33.78
40°	.6560	1.193	3.249	21.28
50°	.5490	.970	2.497	14.01
60°	.4690	.808	1.982	9.83
70°	.4060			
80°	.3560			
90°	.3160			
100°	.2840			

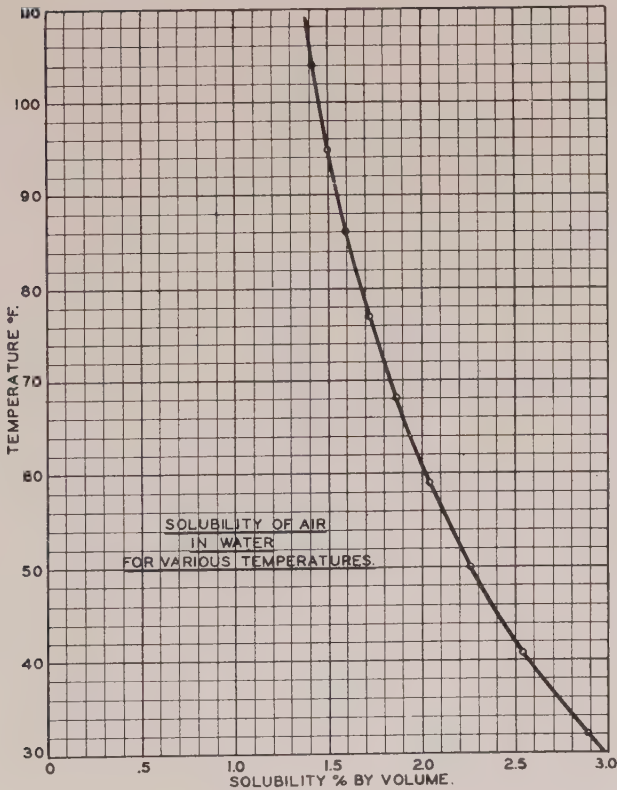


FIG. 1

SOLUBILITY OF AIR IN WATER.*

EXPRESSED AS CUBIC FEET OF DRY AIR AT STANDARD CONDITIONS IN ONE
CUBIC FOOT OF WATER.

(32° F. and 29.92")

° C.	° F.	Solubility	Total Pressure in Atmospheres Air + Water Vapor	Calculated Solubility Where Total Pressure Is One Atmos- phere, Assuming Henry's Law
0.....	32.0	0.02881	1.005	0.02865
1.....	33.8	2808	6	2790
2.....	35.6	2738	7	2715
3.....	37.4	2670	7	2650
4.....	39.2	2606	8	2585
5.....	41.0	2543	9	2520
6.....	42.8	2482	10	2460
7.....	44.6	2424	10	2400
8.....	46.4	2369	11	2345
9.....	48.2	2316	11	2290
10.....	50.0	2264	12	2240
11.....	51.8	2217	13	2190
12.....	53.6	2171	14	2140
13.....	55.4	2127	15	2095
14.....	57.2	2085	16	2055
15.....	59.0	2045	17	2010
16.....	60.8	2005	18	1970
17.....	62.6	1970	19	1935
18.....	64.4	1935	20	1895
19.....	66.2	1901	21	1860
20.....	68.0	1869	23	1825
21.....	69.8	1838	24	1795
22.....	71.6	1808	26	1760
23.....	73.4	1779	27	1735
24.....	75.2	1751	29	1700
25.....	77.0	1724	31	1675
26.....	78.8	1698	33	1640
27.....	80.6	1674	35	1618
28.....	82.4	1650	37	1590
29.....	84.2	1627	39	1565
30.....	86.0	1606	41	1545
35.....	95.0	1503	55	1425
40.....	104.0	1418	72	1325
45.....	113.0	1351	94	1235
50.....	122.0	1297	121	1155
60.....	140.0	1216	196	1015
70.....	158.0	1156	307	883
80.....	176.0	1126	465	768
90.....	194.0	1113	691	658
100.....	212.0	1105	2.000	553

* Winkler, Landolt-Bornstein-Roth Tabellen.

SOLUBILITY OF SODIUM ACETATE.

Temp. C.	Solubility, Grams per 100 Grams of Water	
0		23.5
10		27.3
20		31.5
30		37.2
40		44.5
50		66.0
60		...
70		...
80		...
90		...
100		...

SOLUBILITY OF SOME CHLORIDES.

Temp. C.	Solubility, Grams per 100 Grams of Water						
	CaCl ₂ ^b	KCl ^c	NaCl ^c	MgCl ₂	NH ₄ Cl ^a	ZnCl ₂	BaCl ₂ ^a
0	59.5	28.1	35.7	52.8	29.7	...	30.9
5	61.3	29.7	35.8	...	31.4	...	32.2
10	65.0	31.3	35.8	53.5	33.3	...	33.3
15	68.4	32.9	35.9	...	35.2	79.1	34.5
20	74.5	34.5	36.0	...	37.2	81.2	35.7
25	85.9	36.0	36.1	...	39.3	...	37.0
30	100.8	37.5	36.2	...	41.4	...	38.2
35	107.5	38.9	36.3	...	43.6	...	39.5
40	115.3	40.3	36.5	57.5	45.8	82.2	40.7
45	130.4	41.7	36.6	...	48.0	...	42.2
50	132.6	43.1	36.8	...	50.4	...	43.6
55	134.7	44.5	37.0	...	52.8	...	45.0
60	136.8	45.9	37.3	61.4	55.2	83.5	46.4
65	139.3	47.1	37.5	...	57.7	...	47.9
70	141.6	48.4	37.6	...	60.2	...	49.4
75	143.9	49.6	37.8	...	62.8	...	50.9
80	146.9	50.9	38.1	66.3	65.6	...	52.4
85	149.4	52.3	38.3	...	68.4	...	54.0
90	152.5	53.6	38.6	...	71.3	...	55.6
95	155.8	54.8	38.9	...	74.3	86.1	57.2
100	159.1	56.1	39.2	...	77.3	...	58.8
105	163.2	57.3	39.3	...	80.5	...	...
110	166.0	58.1 *	39.6 **	...	83.8	...	...
115	169.5	...	...	...	87.1	...	...
120	173.2	...	...	...	...	...	...
125	177.0	...	...	...	...	...	...

^a Mulder.^b Roozeboom.^c Berkeley.

* 108 = saturated solution.

** 107 = saturated solution.

SOLUBILITY OF SODIUM AND POTASSIUM NITRATES AND CARBONATES.

Temp. C.	Solubility, Grams per 100 Grams of Water			
	NaNO ₃ ^a	KNO ₃ ^a	Na ₂ CO ₃ ^b	K ₂ CO ₃ ^b
0	73.1	13.1	7.0	89.4
5	76.8	17.7	9.5	104.0
10	80.4	21.9	12.5	109.0
15	84.2	26.4	16.4	110.0
20	88.1	33.0	21.5	112.0
25	92.1	39.8	28.2	113.0
30	96.2	46.8	37.8	114.0
35	100.6	55.6	46.2	115.0
40	105.2	65.5	46.1	117.0
45	109.6	75.1	...	119.0
50	115.1	84.0	...	121.0
55	119.8	93.0	...	124.0
60	124.6	101.1	46.0	127.0
65	130.3	108.8	...	130.0
70	136.0	136.4	...	133.0
75	141.7	153.9	...	137.0
80	148.1	171.1	45.8	140.0
85	154.7	188.1	...	144.0
90	161.3	204.9	...	147.0
95	169.4	226.0	...	151.0
100	177.7	248.6	45.5	156.0
105	185.8	271.0	45.2	161.0
110	194.1	293.6	...	167.0
115	202.3	*311.6	...	173.0
120			...	181.0
125			...	188.0

^a Berkeley.^b Mulder.

* 114° C. B. P. saturated solution.

SOLUBILITY OF SODIUM AND POTASSIUM HYDROXIDE.

Temp. C.	Solubility Grams per 100 Grams of Water		Temp. C.	Solubility Grams per 100 Grams of Water	
	KOH ^a	NaOH ^b		KOH ^a	NaOH ^b
0	97.0	42.0	70		
5		47.5	75		
10	103.0	51.5	80		313.0
15	107.0	63.5	85		
20	112.0	109.0	90		
25			95		
30	126.0	119.0	100	178.0	
35			105		
40		129.0	110		365.0
45			115		
50	140.0	145.0	120		
55			125	213.0	
60		174.0	143	311.7	
65			192		521.0

^a Gerlach.^b Pickering, Ferchland.

SOLUBILITY OF VARIOUS SULPHATES.

Temp. C.	Solubility, Grams per 100 Grams of Water			
	Na_2SO_4^a	K_2SO_4^a	FeSO_4^b	$(\text{NH}_4)_2\text{SO}_4^a$
0	4.67	7.40	15.65	70.6
5	6.73	8.34	18.07	71.8
10	9.08	9.29	20.51	73.0
15	13.54	10.24	23.48	74.2
20	20.41	11.18	26.60	75.4
25	27.89	12.13	29.81	76.7
30	40.95	13.07	32.94	78.0
35	49.09	14.03	36.40	79.5
40	48.20	14.99	40.15	81.0
45	47.41	15.85	44.14	82.7
50	46.72	16.63	48.43	84.4
55	45.96	17.41	53.15	86.2
60	45.24	18.81	54.94	88.0
65	44.65	19.01	54.64	89.9
70	44.10	19.83	50.99	91.6
75	43.60	20.66	47.47	93.4
80	43.28	21.35	43.90	95.3
85	42.97	22.05	40.58	97.2
90	42.66	22.74	37.35	99.2
95	42.46	23.40		101.2
100	42.26	24.06		103.3
105				105.5

^a Mulder.^b Fraenkel.^c Berkeley.

SOLUBILITY OF SUGAR AT VARIOUS TEMPERATURES.

Temp. ° C.	Solubility, Grams per 100 Grams of Water ^a	Temp. ° C.	Solubility, Grams per 100 Grams of Water ^a
0	179.2	55	273.1
5	184.7	60	287.3
10	190.5	65	302.9
15	197.0	70	320.4
20	203.9	75	339.9
25	211.4	80	362.1
30	219.5	85	386.8
35	228.4	90	415.7
40	238.1	95	448.0
45	248.7	100	487.2
50	260.4		

^a Herzfeld.

SPECIFIC GRAVITY OF CALCIUM AND SODIUM ACETATE SOLUTIONS.

Grams in 100 Grams of Solution	Specific Gravity at 17.5° C.	
	$\frac{\text{CH}_3\text{COO}}{\text{CH}_3\text{COO}} \rangle$ ^a Ca	CH_3COONa ^a
1	1.0051	1.005
2	1.0103	1.010
3	1.0155	1.016
4	1.0207	1.021
5	1.0260	1.026
6	1.0313	1.031
7	1.0367	1.036
8	1.0421	1.042
9	1.0475	1.047
10	1.0530	1.052
11	1.0582	1.057
12	1.0634	1.063
13	1.0686	1.068
14	1.0739	1.074
15	1.0792	1.079
16	1.0843	1.085
17	1.0895	1.090
18	1.0947	1.096
19	1.0999	1.101
20	1.1051	1.107
21	1.1105	1.113
22	1.1159	1.119
23	1.1213	1.124
24	1.1267	1.130
25	1.1321	1.136
26	1.1375	1.142
27	1.1430	1.148
28	1.1484	1.154
29	1.1539	1.160
30	1.1594	1.166
31		1.172

^a Gerlach.

SPECIFIC GRAVITIES OF SOME CHLORIDE SOLUTIONS.

Grams in 100 Grams of Solution	NaCl ^a 15 C.	KCl ^a 15 C.	CaCl ₂ ^b 17.9 C.	MgCl ₂ ^a 15 C.	AlCl ₃ ^a 15 C.	ZnCl ₂ ^a 19.5 C.	NH ₄ Cl ^a 15 C.
1	1.00725	1.00650	1.007	1.00844	1.00721	1.010	1.00316
2	1.01450	1.01300		1.01689	1.01443	1.020	1.00632
3	1.02174	1.01950	1.024	1.02533	1.02164	1.029	1.00948
4	1.02899	1.02600		1.03378	1.02885	1.037	1.01264
5	1.03624	1.03250	1.041	1.04222	1.03606	1.045	1.01580
6	1.04366	1.03916		1.05096	1.04353	1.053	1.01880
7	1.05108	1.04582	1.058	1.05970	1.05099	1.063	1.02180
8	1.05851	1.05248		1.06844	1.05845	1.072	1.02481
9	1.06593	1.05914	1.076	1.07718	1.06591	1.082	1.02781
10	1.07335	1.06580		1.08592	1.07337	1.091	1.03081
11	1.08097	1.07271	1.094	1.09495	1.08120	1.100	1.03370
12	1.08859	1.07962		1.10398	1.08902	1.110	1.03658
13	1.09622	1.08652	1.112	1.11300	1.09684	1.120	1.03947
14	1.10384	1.09345		1.12203	1.10466	1.128	1.04325
15	1.11146	1.10036	1.131	1.13106	1.11248	1.137	1.04524
16	1.11938	1.10750		1.14045	1.12073	1.146	1.04805
17	1.12730	1.11465	1.150	1.14084	1.12897	1.155	1.05086
18	1.13523	1.12179		1.15922	1.13721	1.165	1.05367
19	1.14315	1.12849	1.169	1.16861	1.14545	1.175	1.05648
20	1.15107	1.13608		1.17800	1.15370	1.186	1.05929
21	1.15931	1.14348	1.189	1.18787	1.16231	1.196	1.06204
22	1.16755	1.15088		1.19775	1.17092	1.207	1.06479
23	1.17580	1.15828	1.209	1.20762	1.17953	1.218	1.06754
24	1.18404	1.16568		1.21750	1.18815	1.228	1.07029
25	1.19228	1.17234 *	1.229	1.22737	1.19676	1.238	1.07304
26	1.20098			1.23777	1.20584	1.249	1.07575
27			1.250	1.24817	1.21493	1.260	1.07658
28				1.25857	1.22406	1.270	
29			1.272	1.26897	1.23310	1.281	
30				1.27937	1.24219	1.291	
31			1.294	1.29029	1.25184	1.304	
32				1.30121	1.26149	1.316	
33			1.316	1.31213	1.27115	1.329	
34				1.32305	1.28080	1.340	
35			1.338	1.33397	1.29046	1.352	
36					1.30066	1.366	
37			1.361		1.31086	1.380	
38					1.32106	1.392	
39			1.384		1.33126	1.406	
40					1.34146	1.420	
41			1.406		1.35224	1.432	
42						1.446	
43			1.429			1.460	
44						1.473	
45						1.488	
46						1.500	
47						1.518	
48						1.533	
49						1.550	
50						1.566	

^a Gerlach.^b Pickering.

* 24.9.

SPECIFIC GRAVITIES OF SOME CHLORIDE SOLUTIONS (*Continued*).

Grams in 100 Grams of Solution	NaCl ^a 15 C.	KCl ^a 15 C.	CaCl ₂ ^b 17.9 C.	MgCl ₂ ^a 15 C.	Al ₂ Cl ₆ ^a 15 C.	ZnCl ₂ ^a 19.5 C.	NH ₄ Cl ^a 15 C.
51						1.581	
52						1.600	
53						1.615	
54						1.631	
55						1.650	
56						1.669	
57						1.686	
58						1.704	
59						1.724	
60						1.740	

SPECIFIC GRAVITY OF SODIUM AND POTASSIUM CARBONATE SOLUTIONS.

Grams in 100 Grams of Solution	Specific Gravity (15° C.)		Grams in 100 Grams of Solution	Specific Gravity (15° C.)	
	Na ₂ CO ₃ ^a	K ₂ CO ₃ ^a		Na ₂ CO ₃ ^a	K ₂ CO ₃ ^a
1.....	1.01050	1.00914	27.....		1.26787
2.....	1.02101	1.01829	28.....		1.27893
3.....	1.03151	1.02743	29.....		1.28999
4.....	1.04201	1.03658	30.....		1.30105
5.....	1.05255	1.04572	31.....		1.31261
6.....	1.06309	1.05513	32.....		1.32417
7.....	1.07369	1.06454	33.....		1.33573
8.....	1.08430	1.07396	34.....		1.34729
9.....	1.09500	1.08337	35.....		1.35885
10.....	1.10571	1.09278	36.....		1.37082
11.....	1.11655	1.10258	37.....		1.38279
12.....	1.12740	1.11238	38.....		1.39476
13.....	1.13845	1.12219	39.....		1.40673
14.....	1.14950	1.13199	40.....		1.41870
15.....		1.14179	41.....		1.43104
16.....		1.15200	42.....		1.44338
17.....		1.16222	43.....		1.44573
18.....		1.17243	44.....		1.46807
19.....		1.18265	45.....		1.48041
20.....		1.19286	46.....		1.49314
21.....		1.20344	47.....		1.50588
22.....		1.21402	48.....		1.51861
23.....		1.22459	49.....		1.53135
24.....		1.23517	50.....		1.54408
25.....		1.24575	51.....		1.55728
26.....		1.25681	52.....		1.57048

^a Gerlach.

SPECIFIC GRAVITY OF SODIUM AND POTASSIUM NITRATE SOLUTIONS.

Grams in 100 Grams of Solution	Specific Gravity		Grams in 100 Grams of Solution	Specific Gravity	
	NaNO ₃ ^b (20.2° C.)	KNO ₃ ^a (15° C.)		NaNO ₃ ^b (20.2° C.)	KNO ₃ ^a (15° C.)
1.....	1.007	1.00641	26.....	1.190	
2.....	1.013	1.01283	27.....	1.199	
3.....	1.020	1.01924	28.....	1.207	
4.....	1.026	1.02566	29.....	1.215	
5.....	1.033	1.03207	30.....	1.224	
6.....	1.040	1.03870	31.....	1.232	
7.....	1.047	1.04534	32.....	1.241	
8.....	1.054	1.05197	33.....	1.250	
9.....	1.061	1.05861	34.....	1.259	
10.....	1.068	1.06524	35.....	1.268	
11.....	1.075	1.07215	36.....	1.277	
12.....	1.082	1.07905	37.....	1.286	
13.....	1.089	1.08596	38.....	1.296	
14.....	1.096	1.09286	39.....	1.306	
15.....	1.103	1.09977	40.....	1.316	
16.....	1.111	1.10701	41.....	1.326	
17.....	1.118	1.11426	42.....	1.336	
18.....	1.126	1.12150	43.....	1.346	
19.....	1.134	1.12875	44.....	1.356	
20.....	1.142	1.13599	45.....	1.366	
21.....	1.150	1.14361	46.....	1.376	
22.....	1.158		47.....	1.386	
23.....	1.166		48.....	1.397	
24.....	1.174		49.....	1.407	
25.....	1.182		50.....	1.418	

^a Gerlach.^b Schiff.

SPECIFIC GRAVITY OF SODIUM AND POTASSIUM HYDROXIDE SOLUTIONS.

Per Cent. of Solids in Solution	Specific Gravity at 15° C.		Per Cent. of Solids in Solution	Specific Gravity at 15° C.	
	KOH ^a	NaOH ^a		KOH ^a	NaOH ^a
1.....	1.009	1.012	36.....	1.361	1.395
2.....	1.017	1.024	37.....	1.374	1.405
3.....	1.025	1.035	38.....	1.387	1.415
4.....	1.033	1.046	39.....	1.400	1.426
5.....	1.041	1.058	40.....	1.411	1.437
6.....	1.049	1.070	41.....	1.425	1.447
7.....	1.058	1.081	42.....	1.438	1.457
8.....	1.065	1.092	43.....	1.450	1.468
9.....	1.074	1.103	44.....	1.462	1.478
10.....	1.083	1.115	45.....	1.472	1.488
11.....	1.092	1.126	46.....	1.488	1.499
12.....	1.101	1.137	47.....	1.499	1.509
13.....	1.110	1.148	48.....	1.511	1.519
14.....	1.119	1.159	49.....	1.527	1.529
15.....	1.128	1.170	50.....	1.539	1.540
16.....	1.137	1.181	51.....	1.552	1.550
17.....	1.146	1.192	52.....	1.565	1.560
18.....	1.155	1.202	53.....	1.578	1.570
19.....	1.166	1.213	54.....	1.590	1.580
20.....	1.177	1.225	55.....	1.604	1.591
21.....	1.189	1.236	56.....	1.618	1.601
22.....	1.198	1.247	57.....	1.630	1.611
23.....	1.209	1.258	58.....	1.641	1.622
24.....	1.220	1.269	59.....	1.655	1.633
25.....	1.230	1.279	60.....	1.667	1.643
26.....	1.241	1.290	61.....	1.681	1.654
27.....	1.252	1.300	62.....	1.695	1.664
28.....	1.264	1.310	63.....	1.705	1.674
29.....	1.278	1.321	64.....	1.718	1.684
30.....	1.288	1.332	65.....	1.729	1.695
31.....	1.300	1.343	66.....	1.740	1.705
32.....	1.311	1.353	67.....	1.754	1.715
33.....	1.324	1.363	68.....	1.768	1.726
34.....	1.336	1.374	69.....	1.780	1.737
35.....	1.349	1.384	70.....	1.790	1.748

^a Pickering, Mylius and Funk.

SPECIFIC GRAVITIES OF VARIOUS SULPHATE SOLUTIONS

Grams in 100 Grams of Solution	Specific Gravity of Solution at 15° C.			
	$\text{Na}_2\text{SO}_4 \cdot 10\text{H}_2\text{O}$	K_2SO_4^a	$\text{FeSO}_4 \cdot 7\text{H}_2\text{O}^a$	$(\text{NH}_4)_2\text{SO}_4^a$
1	1.004	1.00820	1.005	1.0057
2	1.008	1.01635	1.011	1.0115
3	1.013	1.02450	1.016	1.0172
4	1.016	1.03277	1.021	1.0230
5	1.020	1.04105	1.027	1.0287
6	1.024	1.04947	1.032	1.0345
7	1.028	1.05790	1.037	1.0403
8	1.032	1.06644	1.043	1.0460
9	1.036	1.07499	1.048	1.0518
10	1.040		1.054	1.0575
11	1.044		1.059	1.0632
12	1.047		1.065	1.0690
13	1.052		1.071	1.0747
14	1.056		1.077	1.0805
15	1.060		1.082	1.0862
16	1.064		1.088	1.0920
17	1.069		1.094	1.0977
18	1.073		1.100	1.1035
19	1.077		1.106	1.1092
20	1.082		1.112	1.1149
21	1.086		1.118	1.1207
22	1.090		1.125	1.1265
23	1.094		1.131	1.1323
24	1.098		1.137	1.1381
25	1.103		1.143	1.1439
26	1.107		1.149	1.1496
27	1.111		1.155	1.1554
28	1.116		1.161	1.1612
29	1.120		1.168	1.1670
30	1.125		1.174	1.1724
31			1.180	1.1780
32			1.187	1.1836
33			1.193	1.1892
34			1.200	1.1948
35			1.206	1.2004
36			1.213	1.2060
37			1.219	1.2116
38			1.226	1.2172
39			1.232	1.2228
40			1.239	1.2284
41				1.2343
42				1.2402
43				1.2462
44				1.2522
45				1.2583
46				1.2644
47				1.2705
48				1.2766
49				1.2828
50				1.2890

^a Gerlach.

INDEX

A

Acceleration, 115
Accessories, 231
Acid Cleaning, 188
Active Zones, 134
Air Binding, 175, 238
 Cocks, 200
 Cooling, 146, 147
 Dissolved in Liquor, 131
 In Water, 130, 481
 Leaks, 131, 179, 238, 239
 Pockets, 238, 239
 Pumps, 219
Ammonia Gas Removal, 200
 Pipes, 180, 195
Analysis Triple Effect, 56, 309
Annular Belts, 135
 Space, 135
Appendix, 473
Applications to Specific Industries, 243
Automatic Controls, 203

B

Badger, E. B., & Sons Co., 440
Badger High Speed Evaporator, 440
Badger, W. L., 44, 48, 49, 423
Badger & Shepard, 48, 49
Baffled Calandrias, 139, 140, 412
 Evaporators, 412
Baffles for Entrainment, 101, 114
 In Heaters, 166, 169
 In Steam Belts, 39, 140, 412
Bagasse, 246, 247
Balance, Heat, 56
Barkometer, 12, 479
Barometric Drains, 144, 153, 196
Basic Principles, 51
Basin Heads, 223
Basket Type Evaporators, 352, 443
Batch Concentration, continuous, 215
 System, 142, 215
 Tank, 216
Baumé Scale, 11, 479
Beet Sugar Factory, heat balance in, 293
Bittern, 14
Black Liquor, 359
Boiler, Juice, 296
Boiling Liquids, 34
 Over Condensers, 149

 Over Evaporators, 187
 Point, Raising, 21, 72, 73, 490, 491
 Temperatures, 490, 491
Brick Lined Evaporators, 468, 469
Brix Scale, 11, 479
Buffalo Fdy. & Machine Co., 419, 427, 443
Buflovak Evaporators, 427
Butter Cups, 101
By-passed Condensate, 197

C

Calandrias, 130
 Floating, 444, 445
Calandria Pans, 347
Calcium Chloride, 392, 393
Cane Sugar Factory, heat balance in, 244
Carbonation in Beet Sugar Factory, 301, 312, 324
Carrying Over, 101, 102
Catchalls, 101, 106, 107, 108
Caustic Cleaning, 188
 Soda, 392, 393
 Potash, 392, 393
Centrifugals, 351
Chapman Controls, 209
Charts for Evaporation, 172
Chestnut Extract, 354
Circulation, 109, 119
Circulation Control, 213
 Effect of Level on, 116
 In Evaporator Tubes, 114
 Mechanical, 119
 In Manistee Evaporator, 121
 Power Required for, 124
 Natural, 109
Circulators, 123
Cleaning Evaporators, 186
Cleanliness of Heating Surface, 38
Clearances in Vacuum Pumps, 223
Climbing Film Evaporator, 434
Coefficient of Heat Transmission, Effect of Cleanliness of Surface on, 38
 Effect of Liquor Level, 48
 Effect of Non-condensable Gases, 43
 Effect of Rate of Evaporation on, 41
 Effect of Static Head, 41
 Effect of Temperature of Steam, 39
 Effect of Viscosity, 43

Coil Evaporators, 346, 456, 457, 458
 Heaters, 164
 Pans, 346
 Compound Evaporator, 91, 289, 290, 291
 Compression, Vapor, 93
 Concentration, Batch, 215
 Concentrated Liquor Pump, 195
 Condensate, By-pass, 76, 79
 Pump, 194
 Removal, 179, 193, 238
 Condensers, Counter Current, 148, 149, 150
 Evaporator, 144
 Jet, 144, 217
 Types of, 148
 Condenser Loads, 68
 Condensing Vapors, 33
 Conduction, 24
 Conductivities, Specific, 30, 31
 Conductivity of Surface Films, 32
 Continuous Batch Concentration, 215
 Controls, Feed, 203
 Level, 203
 Control, Pressure, 234
 of Vacuum, 183
 Convection, 24
 Currents, 113
 Cooling Ponds, 156
 Towers, 155
 Correcting Leaks, 179
 Corrosion, 200
 Cost of Steam, 71
 Counter Current Condensers, 148, 149, 150
 Heaters, 160, 169
 Operation, 61, 364
 Covering, 235
 Criss-Cross Evaporator, 133
 Critical Level, 117
 Crystallizers, 348
 Crystallizing Solutions, 345
 Currents, Convection, 113
 Cycles in Cane Sugar Factory, 248

D

Data, 473
 Deformation, 188
 Delicate Materials, 390
 Density, 46
 Control, 183
 Scales, 11, 479
 of Steam, 47
 Tests, 232
 Diffusion in Beet Sugar Factory, 299, 311, 324
 Direct Fired Evaporators, 15
 Distilled Water Cleaning, 187
 Evaporators, 424, 456
 Distillery Slop, 356

Distributed Downtakes, 118
 Downtakes, 113, 114, 118
 Dühring Rule, 23
 Duplex Vacuum Pump, 224
 Dyes and Extracts, 354

E

Early Methods of Making Sugar, 16
 Edwards Pump, 195, 196, 224, 225
 Efficiency of Displacement Vacuum Pumps, 228
 Ejector Condensers, 152, 153, 154
 Entrainment, 101
 Losses, 184, 187
 Entropy, 95
 Essentials for Successful Operation, 237
 European Practice, 294
 Evaporating Apparatus Defined, 51
 Evaporation, 11, 18, 245, 303, 313
 Conditions, Beet Sugar, 295, 296
 Definition of, 11
 of Crystallizing Solutions, 345
 Evaporator Accessories, 231
 Evaporators, Basket Type, 352
 Direct Fired, 16
 Operation of, 171
 Evaporator Charts, 172
 Condensers, 144
 Exhaust Pressure, 293
 Extra Live Steam, 266, 267, 342
 Extracts and Dyes, 354

F

Feed, Hand Control of, 206
 Pipes, 203, 238
 Temperatures, Importance of, 57
 Feeding Systems, 203
 Film Evaporators, 424
 Finishing Pans, 141
 For Paper Mill Waste Liquor, 395
 Flash, 52, 79, 101, 104, 105, 197, 203
 Chambers, 125
 Pots, 226
 Floating Calandrias, 443, 444, 445
 Fluidity of Water, 36
 Foaming, 101
 Foamy Materials, 365
 Forced March, 240
 Fouling, 139
 Foul Surfaces, 186, 189, 237
 Free Power, 16
 Freezing, 13
 Friction Through Calandrias, 130
 Through Vapor Pipes, 96
 Functions of Condensers, 144
 Fundamentals, 237

G

Garrigue Evaporators, 447
 Gas Pockets, 134, 139
 Removal, 200
 Gauges, 231
 Gelatin, 390
 Glue, 390
 Glycerine Evaporation, 394, 447
 Gravity Scale, 12
 Grease Cups, 101
 Greens, 302, 313, 325

H

Hand Control, 206, 207, 208
 Heat Balance, 56, 69
 Example of, 57
 Beet Sugar Factory, 293
 Cane Sugar Factory, 244
 Practical Use of, 69
 From Combustion, 15
 Equivalent of Power, 244
 Heaters, Circulation in, 162
 Counter-current, 169
 Factors Affecting, 160
 Heat Transmission in, 163
 Multipass, 160
 Single Coil Type, 164
 Vertical Multipass, 167
 Single Pass, 166
 Heat Transmission, 24
 By Conduction, 24
 Effect of Circulation on, 111
 Through Composite Walls, 28
 Through Pipes, 25
 Through Surface Films, 29
 Heating, Vapor, 80
 Heavy Group, 392
 High Concentrator, 470
 Extraction, 293
 Speed Evaporator, 440
 Horizontal Tubes, 132
 Horizontal Tube Evaporator, 461
 Horse Shoe Vacuum Pump, 223, 224
 Hughes Separator, 466, 467
 Hydrostatic Head Loss, 41, 73, 115
 Hydrometers, 11, 479
 Tables, 479

I

Inclined Tube Evaporator, 427
 Industries Applications to, 243
 Ingersoll-Rand Co., 227
 Inlets, Steam, 130
 Insulation, 235
 Investigations, 237
 Inverted Syphons, 197, 210

J

Jet Condensers, 144, 217
 Joubert & Goslin, 408
 Juice Boiler, 206
 Preheaters, 283
 Heaters, 245
 Heating, 160

K

Kerr, E. W., 36, 39, 41, 46, 47, 71, 72,
 163, 164, 189, 235, 367, 369, 423.
 Kestner Evaporator, 42, 434
 Paul, 93, 117
 Salt Evaporator, 438
 Vertical Climbing Film Evaporator,
 360, 435, 436, 437
 Kilby Manufacturing Co., 420

L

Lagging, 235
 Latent Heat, 24, 51, 473, 474
 Leaks, Air, 238, 239
 Correcting, 179
 in Tube Sheets, 240
 Level Controls, 117, 203
 Critical, 117
 Gauges, 231
 Lillie Evaporators, 41, 43, 360, 424
 Liquor Circulation, 110, 111
 Pumps, 194, 195
 Live Steam in Beet Sugar Factory, 342
 Long Tube Evaporators, 198, 427, 434,
 445
 Loops, 197
 Loose Grained Castings, 179
 Low Vacuum, 293

M

Manheads, 232
 Manistee Evaporator, 452
 Masse Cuite, 346, 347
 Maximum Power for Vacuum Pumps,
 230
 McAdams, W. H., 32, 34, 44, 45
 Mechanical Circulation, 119, 452
 Cleaning, 188
 Mechanism of Evaporation, 18
 Melada (See Beet Sugar).
 Meter, Steam, 172, 233
 Method of Control, 173, 174, 175
 Metric Equivalents, 475
 Milk Evaporators, 390, 391
 Melted Sugar, 302
 Mollier Diagram, 94
 Moore, Hugh K., 40, 41, 43, 360, 361,
 362, 363, 364
 Multicurrent Heaters, 160

Multipass Heaters, 160, 167, 168, 169
 Multiple Effect Evaporation, 53, 54
 Multiplicity of Downtakes, 70

N

Natural Circulation, 109
 Heat, 15
 Non-Boiling Liquids, 32
 Non-Condensable Gases, 36, 40, 130, 131, 200
 Non-Condensable Gas Removal, 200
 Number of Bodies, selection of, 70

O

Obstructions, 178, 180
 Operating Conditions, Beet Sugar, 295
 Cane Sugar, 249
 Operating Suggestions, 171
 Organic Foamy Materials, 356
 Orifice Tank, 233
 Outlets, Gas, 130
 Outside Settling, 351
 Overflow Control, 211

P

Pan, Vacuum, 303, 313, 325, 346, 347
 Paper Mill Waste Liquors, 359
 Parallel Current Condensers, 157
 Operation, 58
 Parallel Feed Operation, 64
 Pauly, 266, 342
 Percentage of Evaporation, 12, 73
 Performance of Evaporators, 422
 Period of Operation, 172
 Pipes, Non-Condensable, 200
 Vapor, 90
 Pockets, Gas, 134, 139
 Ponds, Cooling, 155
 Spray, 155
 Power for Vacuum Pumps, 228, 230
 Free, 16
 Heat Equivalent of, 244
 Practical Use of Heat Balance, 69
 Prache Bouillon, 93
 Pressure Control, 234
 Double Effect, 337
 Evaporators, 335
 Gauges, 231
 Limits, 71
 Loss in Condensers, 148
 Temperature, 18, 473, 474
 Triple Effect, 338
 Vapor, 18
 Principles, Basic, 51
 Proportions of Tubes, 126

Pumps, Concentrated Liquor, 194, 195
 Condensate, 194
 Purging, 131
 Vacuum, 223
 Water, 218

Q

Quadruple Effects in Beet Sugar, 310
 in Cane Sugar, 256
 Quadruple Vapor Pipes, 137
 Quiescent Zones, 134
 Quintuple Effects in Beet Sugar, 322
 in Cane Sugar, 260

R

Radiation, 24, 235, 245
 Radojet Vacuum Pump, 229
 Raoult's Law, 20
 Rate of Evaporation, 41
 Ratings, 333, 334
 Raw Juice, 301, 312, 324
 Reilly Multicoil Evaporator, 456
 Removal of Condensate, 179
 Reversing, 188
 Rillieux, 141
 Rillieux Evaporator, 417
 Rillieux's First Principle, 53
 Second Principle, 81
 Rotative Dry Vacuum Pump, 225

S

Safety Valves, 232
 Salinometer, 429
 Salt Evaporation, 119, 245, 434, 452
 Traps, 348, 349
 Salting Up, 122
 Saturation, 301, 312
 Saturated Solutions, 12
 Steam, 473, 474
 Vapor, 18
 Scale, 186
 Scaling, 172
 and Fouling, 189
 Schutte and Koerting, 153, 156
 Self Evaporation, 52
 Separators, 101, 106, 107, 108, 467
 Settling, Outside, 357
 Sextuple Effect, Diagrammatic Sketch
 of, 76
 Heat Balance, 76
 Shutling Down, 186
 Sight Glasses, 232
 Single Effect Finishing Pans, 141
 Single Pass Heater, 166
 Slotted Holes in Calandrias, 135

Slop, Distillery, 356
 Soap Lyes, 394, 447
 Solar Ponds, 13
 Solubility of Compounds, 485, 487, 488, 489, 492, 493
 Solutions, 11
 Boiling Points of, 490, 491
 Solution Heaters, 160
 Specific Conductivities, 30, 31
 Gravity of Solutions, 479, 480, 483, 484, 486, 488, 489, 492, 493
 Heat of Solutions, 52
 Heat of Sugar Solutions, 294
 Industries, 243
 Sprays, 14
 Spray Ponds, 155
 Standard Effect, 408
 Static Head Loss (See Hydrostatic Head Loss), 41, 73, 115
 Starting Up, 178
 Steam Belts, 130
 Annular, 136
 Gas Pockets in, 134
 Types of, 130
 Webre, 139
 Consumption in Various Cycles, 67
 Evaporation by, 17
 Injectors, 93, 287, 288
 Jet Vacuum Pumps, 229
 Meters, 172, 233
 Supply from Bagasse, 248
 Tables, 473
 Traps, 193
 Travel through Heating Surface, 132
 Sugar, Beet, 294
 Cane, 244
 Early Methods of Making, 16
 Solubility of, 488
 Solutions, Viscosity of, 482
 Sulphate Liquor, 359
 Sulphite Liquor, 359
 Superheat, 95, 116
 Superheated Gases, 36
 Vapors, 36
 Surface Films, 32
 Sweet Water Condensers, 195
 Pumps, 194
 Removal, 193
 Swenson Evaporator, 418
 Swept Volume of Vacuum Pump Displacement, 229
 Syphons, Inverted, 193, 197, 210
 Syrup, 302

T

Tables, 473
 Tallow Cups, 101
 Tanning Extract, 354
 Temperatures, Boiling, 490, 491

Temperature Control on Evaporators, 16
 Conversion Charts, F., C., R., 476, 477, 478
 Distribution, 73, 74, 75
 Fall, 40, 53, 71
 of Steam, 39
 Pressure, 18, 473, 474
 Terminal Conditions, 71
 Testers, Density, 232
 Thermo-Compressors, 93, 94, 95, 287, 288, 289
 Thermometers, 231
 Thickness of Metal Surfaces, 43
 Thin Juice, 302, 313
 Towers, Cooling, 155
 Transmission of Heat, 24
 Trick Port on Vacuum Pumps, 226
 Triple Effect Cycles, Beet Sugar, 289 to 311
 Triple Effect Cycles, Cane Sugar, 249 to 255
 Heat Balance, 231
 Troubles, 237
 Tube Proportions, 126
 Twaddell Scale, 11, 479
 Twin Vapor Pipes, 136, 137
 Twitchell Lyes, 394
 Types of Condensers, 148
 Evaporators, 407
 Steam Belts, 130

U

U-Pipes, 193, 210
 U. S. Cast Iron Pipe & Foundry Co., 412

V

Vacuum, 124
 Breaker, 232
 Control, 183
 Gauges, 321
 Leaks, 146
 Limits, 70
 Pans, 245, 297, 298, 345, 443
 Pumps, 144, 146, 219
 Pumps, Efficiency of Displacement of, 228
 Test, 178
 Vapor Cells, 284
 Circulation, 112
 Compression, 93, 244
 Heating, 253, 280
 Pipes, 96
 Pressure Relations, 18
 Vaporization, 18
 Velocity Head, 115
 of Circulation, 110, 111, 112
 Venting, 200
 Vent Pipes, 200

Vertical Heaters, 166, 167
Tube Evaporators, 133
Viscosity, 43, 44, 71, 364, 482

W

Walking Beam Pump, 223
Wash Water, 302, 313, 325
Washing Down, 187
Waste Liquor, Paper Mill, 359
Water, Air Contents of, 481
 Hammer, 240
 Heating, 144, 145, 146
 Removal, 238
 Supply for Jet Condensers, 217
 Test, 178
Webre Evaporators, 412
Weiss Condenser, 150

Wellner-Jellinek Evaporator, 419, 420
Westinghouse Condenser, 151
Wet Vacuum Pump, 223
Wheeler, C. H., Manufacturing Co., 157
 158, 229
Wheeler Condenser & Engineering Co.,
 225, 226, 456
Wheeler Evaporator, 457, 458
White Sugar, 293
Worthington Pump & Machinery Corp.,
 224
Wurster & Sanger Evaporator, 394

Z

Zaremba Evaporator, 461
Zig-Zag Evaporator, 138
Zone Concentration, 203

